THE UNIVERSITY OF ALBERTA

COMPUTER CONTROL OF AN EVAPORATOR

by

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The undersigned certify that they have read, and recommend to the Faculty of Graduate Studies for acceptance, a thesis entitled "Computer Control of an Evaporator", submitted by Manfred Fehr in partial fulfilment of the requirements for the degree of Master of Science.

ABSTRACT

A pilot plant scale double effect evaporator was operated under direct digital control using an IBM 1800 Data Acquisition and Control System.

Feed blending was accomplished by the use of flow ratio control. Averaging and cascade control were applied to liquid holdups in both effects.

Product concentration was controlled by a feedback control system which manipulated the steam flow rate. Feedforward control was added to compensate for feed flow and concentration disturbances. In the case of feed flow disturbances, the compensation was based optionally on the feed flow or first effect bottoms flow signals. Only standard direct digital control software was employed to implement the feedforward controller, such that the method used is expected to be generally applicable to similar processes.

An inferential control system was developed and applied that was based on calculated values of product concentration rather than directly measured values.

A series of 22 experiments were conducted to evaluate the performance of the control systems developed. The use of feedforward control alone gave rise to small off-sets in product quality that were eliminated by the addition of feedback correction. The combined controller maintained essentially constant product quality for changes

in feed flow and concentration 20% of their steady state value in magnitude. The inferential control system resulted in slightly better product concentration control than the conventional feedback system due to the fact that the simplified model response actually led, or anticipated, the process response.

Several advantages of a digital computer over conventional analog control instruments were illustrated. They include the possibility of communication between the direct digital control program and user written programs, the ease of implementing various control schemes by interconnection of control loops, the wide range of controller constants available, and the ease with which control configurations may be modified.

ACKNOWLEDGEMENTS

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1 INTRODUCTION

1.1 Digital Process Control

In recent years, digital computers have become commonplace in many industries. They have assumed a variety of duties, such as stock taking, data logging, and process control. Although they are capable of controlling processes directly, without the use of analog controllers, to date only about one quarter of all computers installed in the chemical industry really perform direct digital control. All others are used for applications such as data logging or supervisory setpoint control (4). This situation is likely to change in the future, for the advantages of direct digital control are numerous. Computer control offers more effective use of process information and an almost unrestricted choice and combination of control algorithms. Measurements and setpoints are available in engineering units and may therefore be determined and manipulated with high accuracy. The necessity for human presence is reduced (8).

The rapid increase of direct digital control applications has created a need for standard programs that can be readily adapted to handle a wide variety of control applications. It is now believed (30) that it should ultimately be possible to satisfy ninety percent of all user program requirements with standard off-the-shelf

packages. These standard programs will include direct digital control algorithms, feedforward control algorithms, heat and material balances, adaptive control, dynamic optimization of process behaviour, operator communications, log, scan, and alarm routines. The potential of the software development is illustrated by the fact that in 1966 only fifty percent of software required by users was available in form of ready made packages.

This need for research and development led the University of Alberta to install an IBM 1800 Data Acquisition and Control System in 1967. The installation is available for research in the field of digital process control and as a service facility for other research projects. It has been interfaced with an existing pilot plant scale evaporator in order to demonstrate some of the computer control applications.

1.2 Pilot Plant Evaporator

The double effect concentrating evaporator had been built in this department three years ago. Prior to the installation of the computer, the evaporator was operated with analog controls. It was used for process control studies by Andre (2) and Wilson (31).

Andre designed and supervised the construction of the evaporator. He also tested various possible schemes for controlling the process. He used a forward feed mode of operation and concluded that the product concentration

could best be controlled by manipulating the steam flow. From practical industrial considerations it followed that the feed rate would be determined by requirements elsewhere in the plant and was therefore not available for manipulation. The objective of Andre's work was the development of a mathematical model for the evaporation process.

Wilson redesigned several sections of the unit to improve the operation. They included the recirculation in the second effect, the condensate collection, the vacuum control, and the steam supply. He implemented an analog control unit that adjusted the steam flow via feedforward control in response to feed concentration and feed flow disturbances. He concluded that feedforward compensation represented an important asset of the control system and recommended further work in this direction. After the termination of Wilson's experimental work, the evaporator was moved to a new building. On reassembling the apparatus, the major components remained unchanged, but several modifications designed to improve its performance and versatility were implemented. The feed system was rebuilt to permit blending of individual feed streams. Additional piping was installed that provides for more than twenty different modes of operation. Orifice plates and valve stems were changed to allow for higher throughput rates. With the newly installed process-computer interface, the computer is

able to read all essential process variables and exercise control over them. A switching system was also added to permit the choice between analog control and direct digital control of individual process variables.

1.3 Objectives of the Present Study

The present study was to provide a link between the previous work carried out in this department, and the field of digital computer control. The existing process equipment used in prior experimentation was retained, but the analog controllers were replaced by digital control algorithms.

The specific objectives were to:

a) develop a digital control system for the evaporator

b) illustrate digital control configurations such as ratio and cascade systems

c) implement digital feedforward control

d) demonstrate inferential control

The digital control system for the evaporator was considered in three parts. The first part was the new feed blending system which could be treated as a separate entity. The second part consisted of the evaporation process as such, comprising vacuum, level, and interstage flow controls. The product concentration control formed the third part.

On implementing the digital control system for the evaporator, it was proposed to make effective use of control configurations that involved a network of individual control loops, such as cascade controllers and flow ratio controllers. Their successful application was expected to demonstrate an additional advantage that direct digital control systems offer over analog controls.

Wilson's experimental work had shown that definite improvements in product quality control can be achieved through the use of feedforward compensation for load disturbances. At the same time, the increasing use of digital control systems demands general purpose software for various kinds of duties. Both these factors taken together made it desirable to develop a digital feedforward controller that would be assembled solely from available smaller pieces of standard software and would therefore be applicable to most processes of a similar nature.

In certain industrial applications, it is desirable to exercise control over a process variable without having to measure it physically (24), (14). Mathematical simulation of the process presents an alternate method for obtaining the value of such a variable. It was proposed to apply this type of control to the evaporation process. It would consist of a mathematical model of the process that would be given values of process inputs, and calculate the value of the product concentration at fixed intervals.

This calculated, or inferred, value would be supplied to the digital feedback controller in place of the measured value.

1.4 Experimental Work

The experimental program for this study was designed to demonstrate the response of the evaporation system to changes in feed flow and concentration when operated under conventional feedback, feedforward, combined feedback-feedforward, or inferred feedback control.

A series of preliminary experiments was also necessary to determine satisfactory steady state throughput rates, to calibrate all instruments, to verify the effectiveness of the proposed control schemes, and to demonstrate satisfactory performance of the computer control system.

During the experimental program, the analog control units were only used for recording of variables. To control the operation, the digital control facilities were used throughout. Start-up and shut-down were facilitated by a set of output stations that permitted manual setting of control valves.

2 LITERATURE SURVEY

2.1 Evaporation Theory

Both Andre (2) and Wilson (31) have treated the theory of evaporation in detail. Andre has placed special emphasis on the aspect of heat transfer in evaporators and the possibilities of mathematical modelling. He developed a dynamic model that adequately represented the evaporation process. Wilson's report is oriented towards the operational aspects of evaporation equipment. He used a steady state material balance to provide the mathematical basis for his feedforward control unit.

In a recent article by Varcop (28), equations have been derived that describe the dynamic behaviour of boiling liquid in the tubes of a forced circulation evaporator. The special dimension of this treatment is that it considers pressure variations along the tubes.

Various types of level control systems for vessels containing liquid have been described by Buckley (7). Among them is the averaging control system that has been applied to both effects of the evaporator in this study.

2.2 Data Acquisition and Control Computers

The general field of computers and process control is reviewed annually in Industrial & Engineering Chemistry (29). General trends of computer applications are discussed,

and articles published on this subject over a one-year period are listed.

A very practically oriented discusssion of the problems that industrial computer applications present has been given by Control Engineering (34). The series provides information on all aspects of computer usage for process control, such as software, hardware, software-hardware interaction, various kinds of languages used in the development of software, as well as routine type programming techniques. Despite the continuing trend towards standardization of process control software, emphasis is repeatedly placed on the necessity for the engineer to be familiar with the operation of the computer in order to be able to make full use of its capabilities.

A report on a compressor station controlled by a digital computer has been published by Friedli (11). It contains a qualitative description of the functions assumed by the computer.

In a process control environment, it is generally necessary to assign different levels of priority to different types of programs. This ensures that process monitoring is not delayed or interrupted by non-process computer users. The concept of interrupts based on priority assignments has been treated by Gelder (13).

2.3 Computer - Process Interface

Articles can be found that describe the techniques employed in converting digital signals into analog signals and vice versa. This signal conversion is an essential part of the computer - process interfacing equipment. Analog-to-digital signal conversion has been treated by Daley (9), and digital-to-analog conversion by Harris (15). An article by Abernethy contains information on multiplexers (1).

2.4 Control Software

In the development of control software, vendor and user usually complement each other. The computer vendor supplies the basic programs required for a particular type of application. The user supplements them with special programs that suit his own specific purpose. In the present case, the main program representing the implementation of the direct digital control system has been supplied and described by the vendor (22). Additional descriptive material as well as supplementary programs are being added as new applications arise (35).

A different application of the same computing system as that used in the present study has been described by Butler and Merryweather (8). The process in question is a glass fibre plant that is being controlled by single variable digital feedback controllers. It is stated that more elaborate modes of control, such as cascading of several loops,

will be added to the system in the future. The application of the direct digital control system to a pilot plant scale evaporator, and the addition of feedforward and inferential controllers to the feedback controls, mark the present work as an additional step following the study by Butler and Merryweather.

Bakke (5) has developed an algorithm that can be used in addition to multiple mode digital control to provide improved regulation for processes with inherent dead time periods. Implementation involves communication between the digital controller and a user written program at every sampling instant. This need for regular communication also arose in the present study on implementing the inferential controller.

2.5 Feedforward Control

A classical definition of feedforward control is contained in an article by Luhrs (21). It is a control action that converts data on process disturbances into corrective action before actual changes in the controlled variable occur. Thus feedforward is a control acting on news rather than history. The principle of feedforward control has been used in the power industry for the last forty years. Since the advent of computers, it has gained wider acceptance in the industry generally.

The derivation of a feedforward transfer function for a process with one disturbance variable and one controll-

ed output variable has been presented by Luecke and McGuire (20). The first part of this article contains the direct analog to the method used in the present study to determine the relationship between the steam flow rate and feed concentration disturbances. The authors continue from there to provide compensation for model errors and to include provisions for limiting process conditions.

A general method of designing feedforward controllers for processes described by dynamic time-invariant linear models has been presented by Bollinger and Lamb (6). The matrix approach is used to represent the process and the controller mathematically. The study is theoretical, no experimental application having been attempted.

Feedforward control achieved with the use of analog components has been described by Wilson (31) and Michon (23). Wilson's experience with the evaporation process has been a point of departure for the present study. Michon has reported on the successful implementation of feedforward control in electrolytic chlorine manufacture. In this case, the pressure of chlorine vapour downstream of the electrolytic cells was regulated by using information on the condition of the cells, such as current density and liquid flow rates.

A recent publication of the Foxboro Company describes various applications of feedforward control as implemented with analog instrumentation (12). A feedforward control system for a double effect evaporator is presented in

which the feed tank level signal generates a series of control actions designed to adjust both steam and feed liquor flow rates.

Reports by Woodley (33) and Tierney et al (26) describe applications of digital feedforward control. The object of Woodley's work was to regulate the temperature of the liquid leaving a heat exchanger. The feedforward controller was activated by disturbances in the liquid flow entering the exchanger. In his implementation, Woodley called for feedback and feedforward calculations at specified intervals and combined the results. The concept is the same as that used in the combined feedforward - feedback controller for the evaporator in the present study.

In the work of Tierney et al, the product concentration of a stirred tank reactor was controlled. Disturbances in feed temperature and feed concentration were considered. Controller equations, even for the feedback controller, had to be programmed especially for this process. For the present evaporation study, the use of the standard direct digital control package has eliminated this need, as the package provides algorithms for different modes of control, and only the values of the controller constants have to be supplied.

2.6 Inferential Control

From the literature reviewed, it becomes apparent

that the idea of inferential control has found acceptance in a variety of industries.

Schaefer has described this type of control as applied to a lime stone quarry (24). The raw material is excavated at different locations and dumped onto a conveyor system via various crushers. The quality of the material at all excavating stations is observed frequently and entered as data into the computer. The combined quality of all material reaching the point of convergence is then calculated. A feedback signal system prevents input from points with off-specification material until such time that material from other input points balances the off-set. The study of the evaporation process followed essentially the same concept, with the additional feature that eventually, troublesome measuring devices can be dispensed with.

Inferential control has also been treated by Harbaugh (14). Although the application is different from that of the present study, an interesting parallel can be drawn, as the concept and the implementation are the same. The system described is a steel rolling mill, i.e. a purely mechanical operation. Input variables are measured and their values transmitted to the computer. The computer will use the mathematical process model to calculate the corresponding product quality. This calculated product quality is compared to the setpoint and the error used for control action.

A textbook approach to inferential control has been undertaken by Kelley (19). He presents the mechanism producing this type of control as a Ziebolz controller, which he defines as an automatic regulating device whose control action is based on information obtained from a simulation of the process under study. The main difference between the closed loop feedback controller and the Ziebolz or inferential controller is that the former is activated by instantaneous error functions, whereas the latter derives its control action from predictions obtained from a mathematical model. Kelley insists on the fact that the Ziebolz controller is more effective than a conventional feedback controller because it works with the future of the variable under control, instead of with its present or past. The problem of controlling a process by this method is essentially reduced to a problem of simulation. Kelley's treatment remains qualitative, as no specific experiments are presented. It can be regarded as a starting point for future work on this subject, which is, as the author states, still in the infancy of its development.

Several industrial organizations are using or developing control schemes similar to those used in this work. However, for competitve reasons most of this work has not been published or is reported in only very qualitative terms.

3 PROCESS EQUIPMENT AND CONTROLS

3.1 Introduction

The double effect evaporator that formed the experimental apparatus is represented schematically on Figures 3.1 and 3.2. The detailed equipment configuration is shown on Figure B-19 of Appendix B. This final configuration has evolved gradually over the past two years, in a continuing effort to extend the capability and flexibility of the equipment. The extent of the modifications can be appreciated by comparing Figure B-19 to the flow sheet of the same evaporator as used by Wilson (31). Apart from the changes noticeable on the figure, the equipment has been moved to a new building and interfaced with the computer. With the completed interface, all essential process variables can be regulated on an optional basis either under local analog, analog supervisory, or under direct digital control. During the present study, the analog controllers were only used in some preliminary runs. The equipment as assembled in the new building is described in the operating manual available in this department (32).

It was also decided to abandon the use of sugar solutions because of their objectionable sticky properties and their tendency to ferment. All experiments were conducted with triethylene glycol solutions. Triethylene glycol is a colorless liquid that meets all the safety criteria

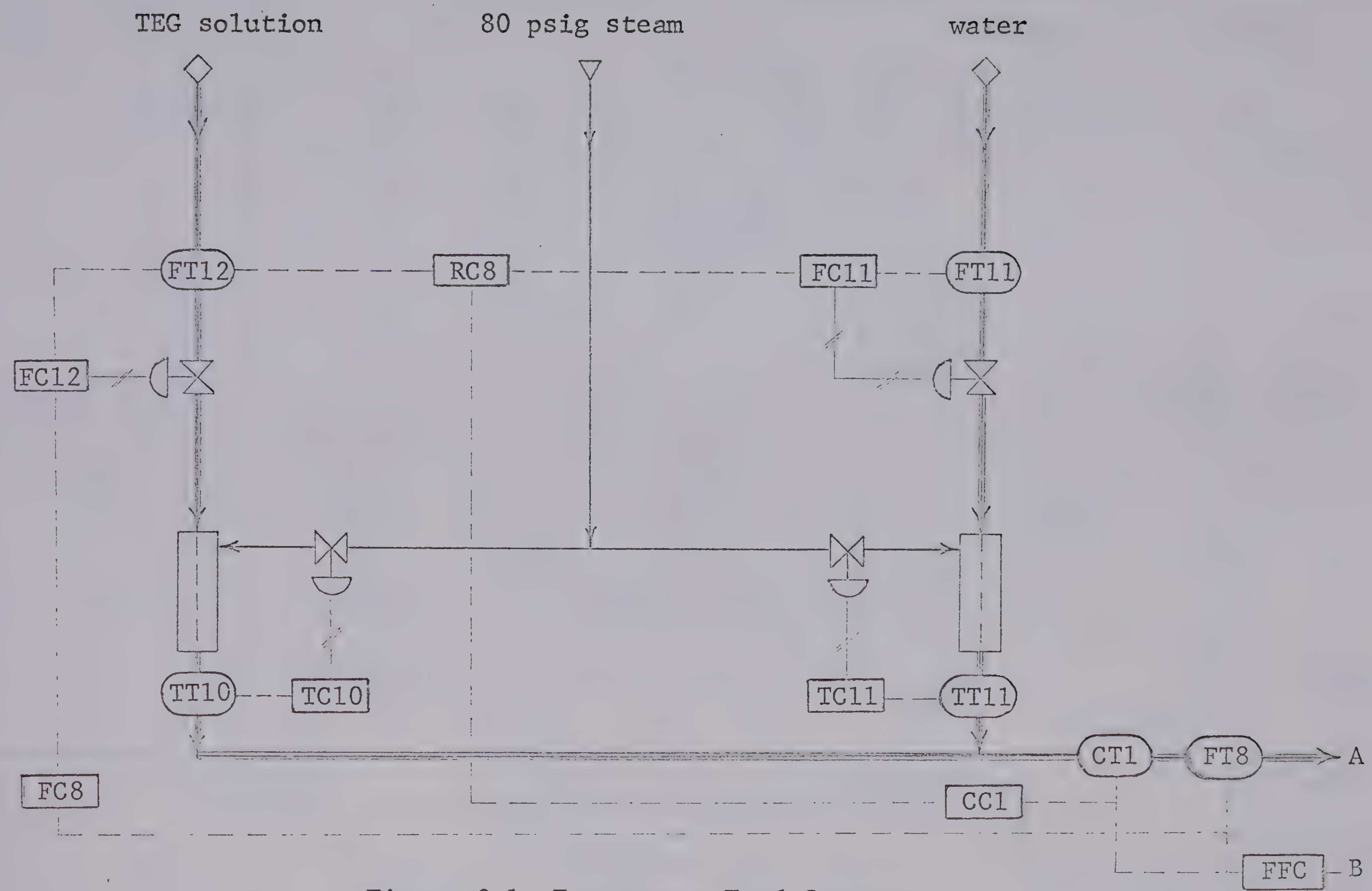


Figure 3.1 Evaporator Feed System

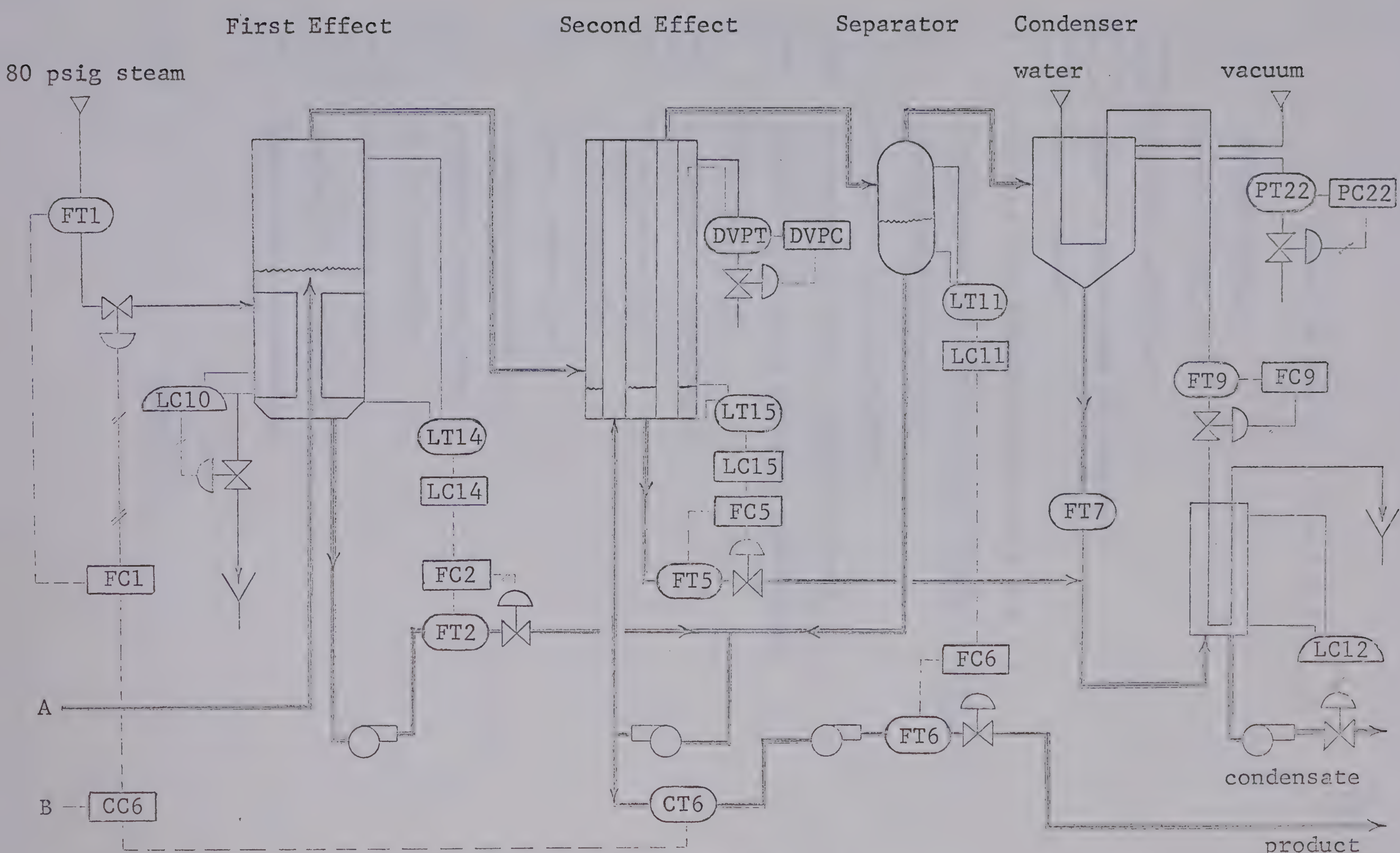


Figure 3.2 Double Effect Evaporator

Legend to Figures 3.1 and 3.2

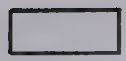
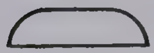
◇	process stream
▽	service stream
	digital controller
	signal transmitted to computer
	local control
A	process link between Figures 3.1 and 3.2
B	instrumentation link between Figures 3.1 and 3.2
CC	concentration controller
CT	concentration transmitter
DVPC	differential vapor pressure controller
DVPT	differential vapor pressure transmitter
FC	flow controller
FFC	feedforward compensation
FT	flow transmitter
LC	level controller
LT	level transmitter
PC	pressure controller
PT	pressure transmitter
RC	ratio controller
TEG	triethylene glycol
TC	temperature controller
TT	temperature transmitter

Table 3.1

to be considered in a student laboratory environment. Its volatility is very much less than that of water, and it does not exhibit a boiling point rise in the temperature and concentration ranges considered (27).

3.2 Design Changes

3.2.1 Feed System

The evaporator feed system had been completely redesigned when the apparatus was reassembled in the new building. As had been pointed out by Andre (2), in industrial installations the feed rate to an evaporator is usually determined by conditions prevailing elsewhere in the plant, and is therefore not available for manipulation. The feed system as used in the present study is shown as a separate unit in Figure 3.1. The network of controllers that has been successfully employed illustrates the flexibility of the direct digital control system.

The equipment configuration and controls shown allow for independent changes in concentration, flow rate, and temperature to be introduced. Seven controllers were required to regulate these three variables. Detailed descriptions of the individual digital controllers are contained in Appendix D.

The feed consisted of two streams: triethylene glycol solution (F 12) and water (F 11) that could be

blended in such proportions as to yield the desired concentration (C 1) of the total feed flow (F 8). The solution flow controller received its setpoint from the total flow controller. The ratio controller used the measurement of the solution flow and the ratio constant supplied by the concentration controller to determine the setpoint for the water flow. Setpoints for the total feed flow rate and concentration were entered by the operator. Provision was also made for operating the concentration controller manually in cases where the measurement obtained from the refractometer was found faulty. Heat exchangers were available to control the temperature of the two individual streams. Satisfactory operation was achieved with a constant feed temperature of 195°F, which ensured the prevention of boiling in the feed piping even under severe load disturbances.

The flow controllers were first tuned individually during the testing period. The control constants and filters were later adjusted as need arose under actual operating conditions, such that disturbances could be introduced without resultant cycling. As the feed temperature setpoint was kept constant at all times, the two temperature control loops had to respond only to load changes resulting from steps in flow rate. Over the extended experimental period, the momentary temperature disturbance was kept to ± 4°F by the two temperature controllers. A list of controller constants is contained in Appendix D.

3.2.2 Evaporation System

On reassembling the apparatus, provisions were made for operating the evaporator in at least twenty different modes. Although this particular feature did not affect the present study, it made the unit adaptable for various kinds of investigations.

The closed loop operation that became feasible with the new feed system involved recirculation of the product to the feed tanks. In order to avoid elevated liquid temperatures in the tanks, a cooler was designed and installed that cooled the product to room temperature.

To reduce the noise in the recordings of the bottoms flows from both effects, the orifices were repositioned so that they are now located upstream of the flow control valves.

It was also felt that the evaporator in its present condition could handle much higher throughput rates than those used by Andre (2) and Wilson (31). Consequently, the orifice plates and valve stems in nearly all lines were changed to allow for higher liquid and steam flow rates. The equipment was then operated at an increased capacity, which created conditions more typical of industrial practice. It was also expected that the increased rates would reduce the total hold-up time of the system. Appendix B contains a list of orifice diameters and valve stem sizes as well as calibration curves for all recorders.

Other new installations included a drum in the steam supply line to dampen out pressure fluctuations in the header, cartridge filters in the recirculation lines from the feed pumps to separate accumulated solid particles, and additional piping in the product line to permit pumping of product to three different tanks.

3.3 Evaporator Controls

3.3.1 Averaging Level Control

The control system for the evaporation process is shown on Figure 3.2. In the case of liquid levels, at least two choices of control criteria are available. Firstly, a level can be controlled at a constant value. The outlet flow from the vessel will then have to vary in the same manner as the inlet flow. Secondly, the level can be allowed to rise or fall within the physical limits of the process equipment to absorb disturbances in the inlet flow and only slowly transmit them to the outlet flow. The liquid levels in both effects of the evaporator were controlled by manipulating the outlet flows.

Simulation studies carried out in this department had shown that the product concentration is affected to a considerable extent by sudden changes in the interstage flow rates (10). It was therefore decided to use averaging level control of the type described by Buckley (7). It consists of a level control loop with proportional band equal to the

allowable total fluctuation in liquid level and no or very weak integral action. With this type of control, the outlet valve will be completely shut when the level is at the lower limit, and completely open when it is at the upper limit. Buckley's mathematical treatment shows that maximum regulation of outflow against inflow or pressure disturbances can be achieved with a cascade level-flow control system. Since in the present case, outflow regulation was critical to product concentration control, both level controllers were cascaded to flow control loops, which resulted in satisfactory smoothing of outlet flows. Actual response curves of both levels and bottoms flows resulting from feed flow disturbances are presented in Appendix D.

In the earlier stages of the experimentation, the first effect liquid level control presented difficulties the real cause of which was only found much later in the program. The sightglass used for level readings was connected to the vessel by quarter inch tubing. Apparently, vapour would condense and collect in this tubing, creating a pressure difference between the vessel and the sightglass. As this pressure difference reached a certain value, the water column was blown out of the tubing, and the resulting sudden pressure increase in the sightglass caused the indicated liquid level to fall instantaneously by approximately 1½ inches. This sequence of events was repeated every thirty to thirty-five minutes. Once the cause was apparent,

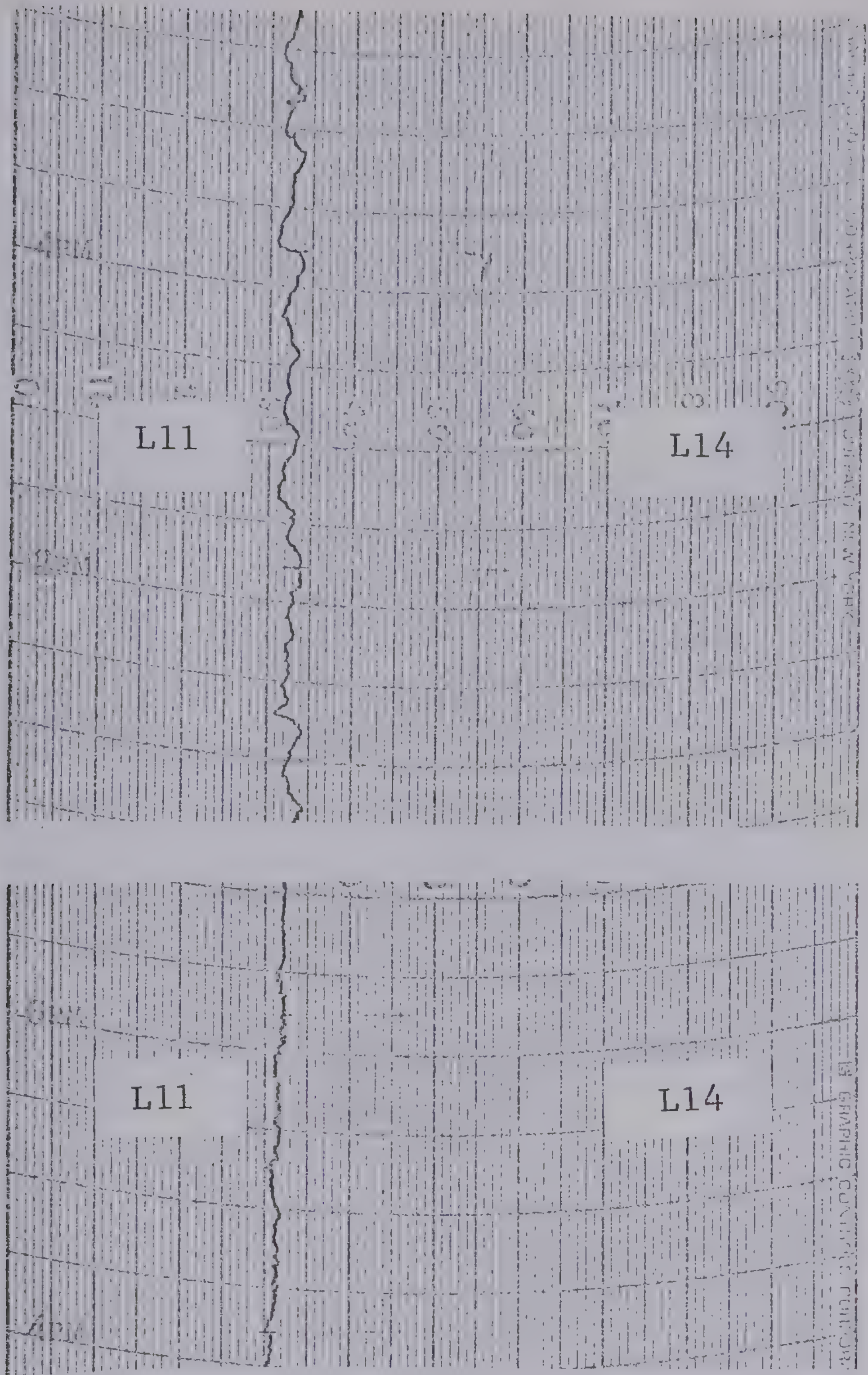
the sightglass was removed, and the level transmitter connected directly to the top and the bottom of the vessel. Although the resulting signal was slightly noisier, the improvement in level control was obvious, as may be seen on Figure 3.3.

3.3.2 Basic Controls

The vacuum in the second effect and the cooling water flow to the condenser were both regulated by simple feedback loops. A differential vapour pressure controller was available to remove accumulated air from the second effect steam chest. The condensate from the first effect steam chest and from the rundown tank was removed by local level controls.

A cascade controller admitted the condensate from the second effect steam chest to the rundown tank. In the original installation, the orifice was located downstream of the control valve. The pressure in the chest ran at 4 to 5 psig, while the rundown tank was kept under 15 in Hg vacuum. This relatively high pressure drop caused severe flashing of the almost saturated liquid that made flow recordings noisy. The valve and the orifice were interchanged, such as to place the orifice upstream of the valve. This arrangement resulted in considerably improved flow recordings, as is illustrated on Figure 3.4.

Another problem arose from fluctuations in the vacuum header. The vacuum service in the building was main-

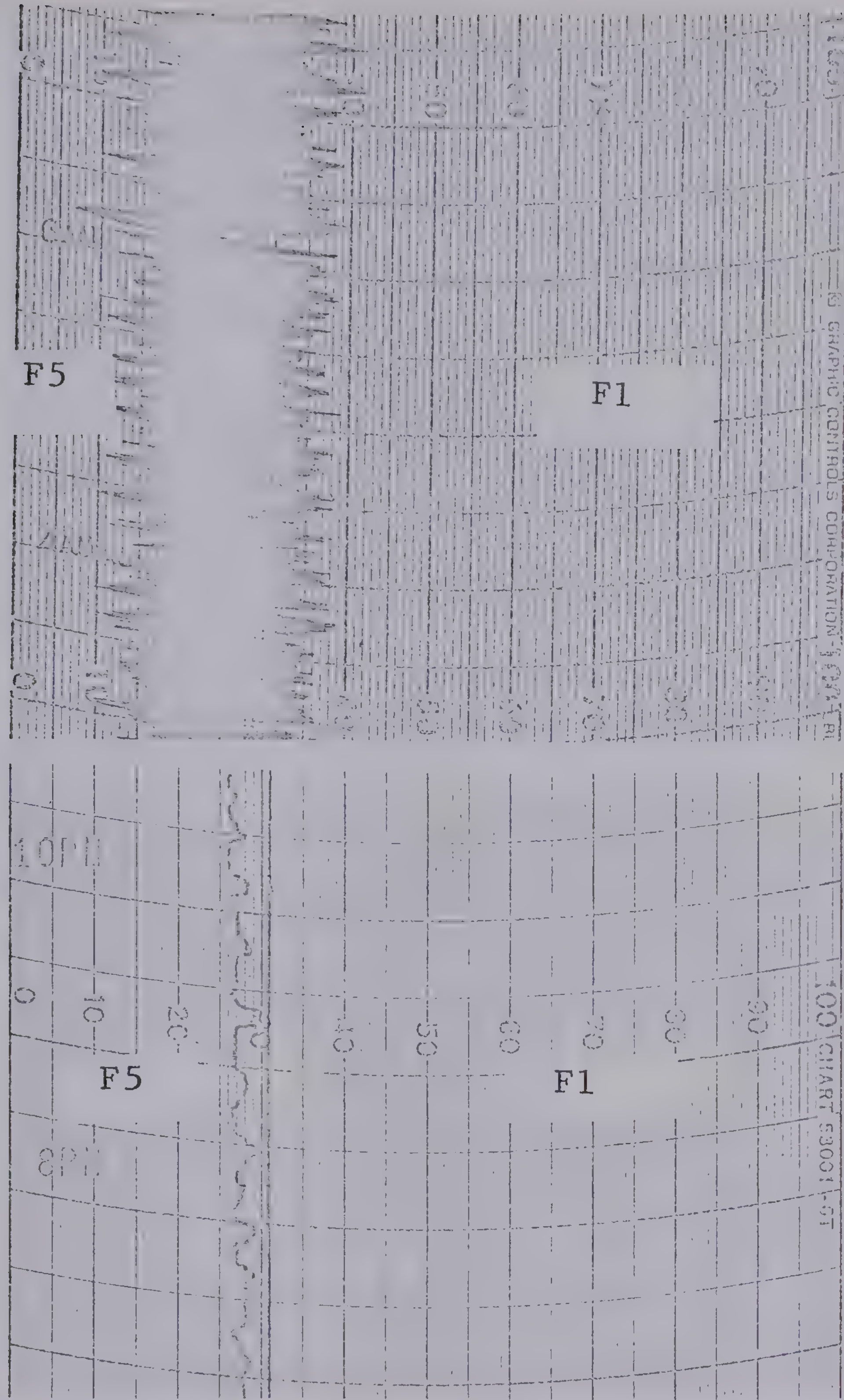


Recordings of Liquid Levels

above: with sight glass on L14

below: without sight glass on L14

Figure 3.3

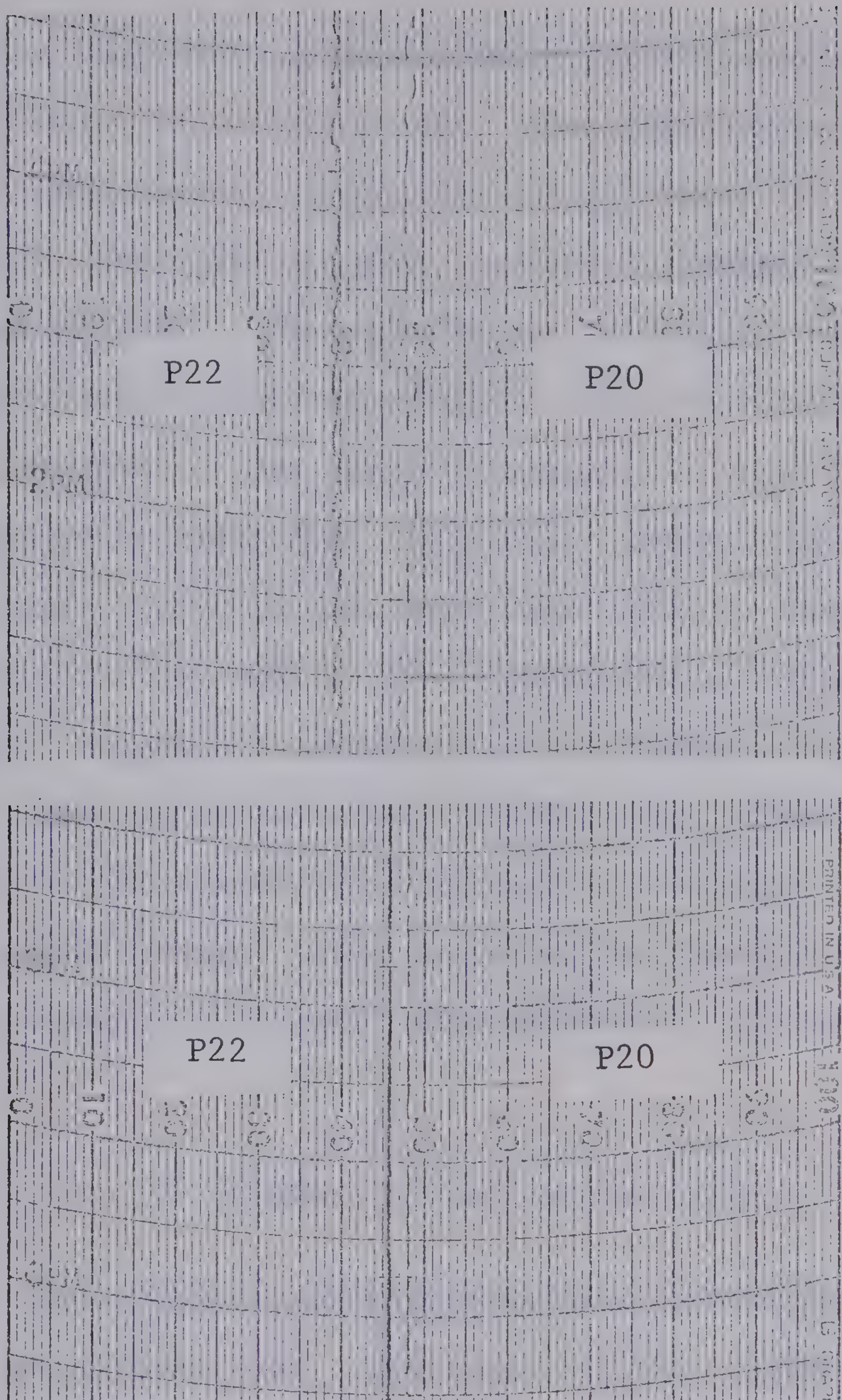


Recordings of Condensate Flow F5

above: orifice downstream of control valve

below: orifice upstream of control valve

Figure 3.4



Recordings of Vacuum P22

above: without pulsation damper

below: with pulsation damper

Figure 3.5

tained by a pump that operated under on-off control, supplying a vacuum of 20 to 24 in Hg. The fluctuations caused by this pump affected the operation of the plant to an extent that had not been anticipated, as they propagated through the entire process. The problem was solved by constructing a pulsation damper and inserting it between the vacuum header and the condenser. The damper consisted of the 6.1 ft^3 steam condensate tank with 15 feet of quarter inch tubing on each side. The tank was made available by leading the condensate from the first effect steam chest directly to the sewer. Figure 3.5 illustrates the results achieved.

The values of several additional variables were available to the computer to be used in material and energy balance calculations. They included the condensate flow from the condenser and the temperatures of all essential streams.

3.3.3 Product Concentration Control

The product concentration was controlled by manipulating the steam flow. The basis of this control system was a cascade concentration-flow controller as shown on Figure 3.2. Additional features of the concentration control system, such as feedforward compensation and inferential control, will be described in Chapter 5.

The constants for the feedback controller were determined empirically during a test run. Values reported

by Wilson (31) were used as a starting point. The preliminary experiment showed that these constants gave rise to an extended period of oscillations in response to a step type load disturbance. It was possible to reduce the oscillations considerably by decreasing the gain. The reset time was then increased to speed up the elimination of offsets.

3.4 General Operating Description

At the outset of the experimental program, a steady state of reference was established that was used as a pivot for all experiments. A set of data for this reference state is presented in Appendix A. For all experiments conducted, a forward feed mode of evaporator operation was employed. Reference is made to Figures 3.1 and 3.2 for the following description.

The feedflow consisted of two merging streams, water and triethylene glycol solution. Mixing took place in the pipe between the point of convergence and the first effect. The proper feed concentration was established by manipulating the ratio of the two flow rates. The two feed streams were heated individually by passing them through steam jacketed heat exchangers. The vapour produced in the first effect was condensed in the steam chest of the second effect and its condensate admitted to the condensate collection system. The bottoms flow from the first effect reached

the second effect at the suction side of the circulation pump. The liquid in the second effect was circulated through the vertical tubes at a velocity of 4.2 ft/sec. Product was withdrawn at the discharge side of the circulation pump and pumped to storage tanks via a cooler. The vapour produced in the second effect was condensed in the condenser and admitted to the rundown tank whose level was kept constant by pumping the condensate back to the water storage tank. The first effect consisted of a calandria type heat exchanger with a total surface area of 9.88 ft^2. The second effect was a vertical tube exchanger with a surface of 4.1 ft^2.

More detailed descriptions of the equipment can be found in references (32), (31), and (2).

4 COMPUTER SYSTEM

4.1 Computer Equipment

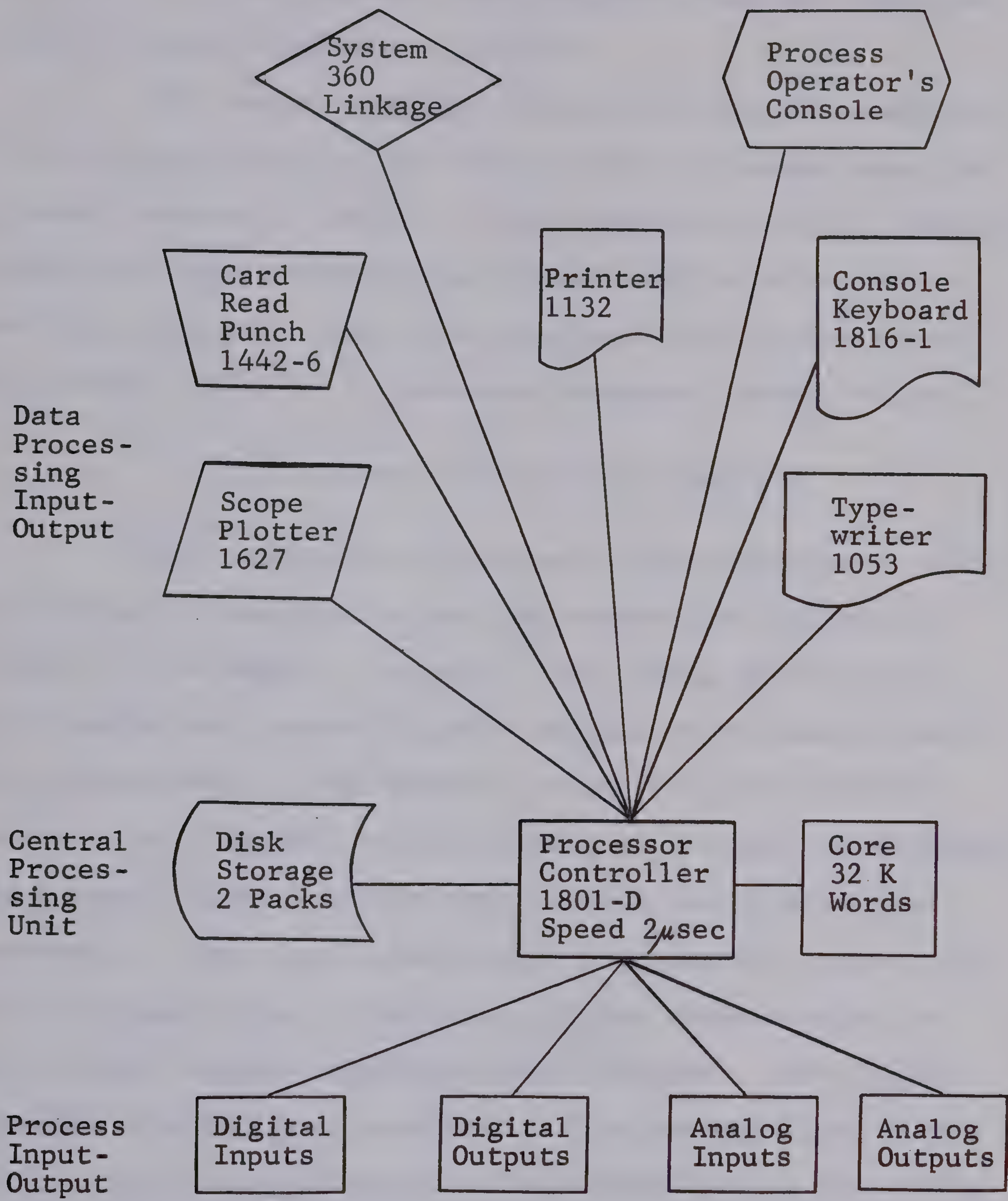
4.1.1 Summary of System Components

The data acquisition and control system that was available for this work is represented schematically on Figure 4.1. Apart from the central processing unit, the components most vital to this study were the analog inputs, the analog outputs, and the process operator's console. The card read-punch unit, the printer, and the plotter were used for purposes of data processing.

4.1.2 Process Operator's Console

The process operator's console consisted of a panel, a typewriter, and a keyboard. The panel contained a set of push buttons that facilitated operator communication with the computer. Actual data input and output proceeded via the typewriter-keyboard unit.

The functions of the process operator's console can be summed up by stating that it presented the physical means of communication between operator, computer, and process. More specifically, it served to select and address process variable records, to display status information on process variables, and to supply or change parameters used for controlling or recording process variables. Process variable records will be discussed in Section 4.3.3. The



IBM Data Acquisition and Control System

Figure 4.1

execution of disk resident user programs could be initiated from the process operator's console.

All analog signals transmitted from the computer to the process had to pass output stations located near the process operator's console. These stations could be used to break the signal transmission circuit and to select valve settings manually. They also displayed the control signal, in percent, going to all computer operated control valves.

4.1.3 Process Input-Output Handling

The functions of the analog input equipment is the collection of analog data and its conversion for presentation to the digital computer. The analog data in this application are current signals emitted by standard industrial transducers. The hardware to convert this current signal into a digital value consists of a signal conditioner, an analog filter, a multiplexer, and an analog-to-digital converter. The signal conditioner converts the current into a voltage signal. The analog filter reduces noise in the voltage signal arriving at the converter. The flying capacitor multiplexer contained in the analog filter circuit allows voltage signals to be transmitted to the analog-to-digital converter. Since this converter operates at high speeds, it can be time-shared among a number of input voltages through an analog relay multiplexer. With this system, the computer can read up to one hundred analog inputs per second. The analog-to-digital converter used allows for a fourteen

Sequence of Signal Conversions

Device	Output Signal
variable measured, e.g. flow	lbs/min
measuring device, e.g. orifice	in. H20
transducer, e.g. dp-cell	10 - 50 ma
signal conditioner, e.g. resistor	1 - 5 volts
analog filter time constant 0.33 sec	1 - 5 volts
signal converter analog-to-digital (14 bits)	6553-32767
program manipulation e.g. range expansion	0 - 32767
digital calculations	0 - 32767
signal converter digital(10 bits)-to-analog	0 - 10 volts
current output station, facility for reading and manipulation	4 - 20 ma
current-to-air converter	3 - 15 psig
final control element e.g. valve	lbs/min

Table 4.1

bit resolution.

The function of the analog output equipment is the conversion of digital values into current signals acceptable to analog instruments such as current-to-air converters. Digital data is stored in output channel registers connected to digital-to-analog converters. The digital values are converted into voltages with a resolution of ten bits. The voltages so obtained are amplified and then converted into 4-20 ma current signals by the current output stations.

A typical sequence of signal conversions is shown on Table 4.1. Detailed descriptions of the equipment involved are contained in publications of IBM Corporation (17, 18).

4.2 Operating System

4.2.1 Time Sharing Executive System

There are two distinctive features of real-time computer applications as compared to all other applications. Firstly, a continuous need exists for communication between the computer and the process under control. Secondly, the computing system has to keep pace with the process, i.e. it has to update records, measurements, and output signals at regular intervals. The Time Sharing Executive System is a group of programs that permit the control of processes while at the same time providing an off-line monitor system for data processing and scientific computations (17, 18).

The framework of the Time Sharing Executive System

is provided by the executive program which resides permanantly in core storage. It consists of the most important programs, sub-routines, and storage areas that are essential for real-time applications. Most other portions of the system are stored on disk and are only brought into core when required for specific functions.

Since the process under control will not utilize all of the computer's time and resources, a certain amount of background duties can be performed by the computer while monitoring the process. The notion of time sharing implies the sharing of computer time, storage, and input-output facilities among on-line and off-line programs or users.

The core storage size is not sufficient to contain all instructions required for the execution of all functions at any one time. It is therefore necessary to break down the total set of instructions into smaller pieces that reside on disk and are available for immediate transfer to core storage when required. These smaller pieces of a set of instructions are stored on disk in executable core image format and are referred to as core loads.

The Time Sharing Executive System consists essentially of two groups of programs; namely, control programs, and processing programs. In general, control programs supervise the execution of processing programs and user programs. The three control programs within this system are the Temporary Assembled Skeleton, the System Director, and the Nonprocess Supervisor. The System Director is the nucleus

of the skeleton executive. It exercises control over all operations connected with real-time operation. It handles the transfer of control between on-line and off-line operations. It governs the transfer from one user specified core load to the next. This function was important to the implementation of the inferential control system described in Chapter 5, where a number of core loads had to be executed in a given sequence. The System Director recognizes, controls, and directs the servicing of interrupts. Interrupts are an essential feature of an on-line real-time control system operated on a time sharing basis. Different priority levels are assigned to all types of programs according to the relative importance that is attached to the time of their execution. The priority scheme thus established basically causes a given series of programs to be executed selectively, rather than sequentially. With respect to the present study, it was important that duties related to the control of the process would be performed by the computer with preference over other duties to prevent disruption of the process.

The System Director also supervises the operation of interval timers. These clocking devices allow for a given program to be executed at fixed intervals, a feature that has been used to advantage in the implementation of the inferential controller.

4.2.2 Programming Considerations

As the skeleton executive occupies a considerable part of core storage, the area available for the execution of user programs, referred to as variable core, was restricted. This factor had to be considered when the special programs needed for this study were written. In general, the programs were divided into small logical units that could be stored as core loads and executed in the proper sequence.

Since there was only a limited number of priority levels available on this system, many process related programs shared the same level. These included the program monitoring the process operator's console. During the time of the experimental program, efficient coordination of various operations involving these programs became imperative. As an example, it was not possible to use the process operator's console while graphs were being produced on the plotter or material balance calculations were being performed.

Facilities were available to accumulate process data in files established in the disk storage unit. With this feature, graphs of process variables as functions of time could be produced on the system plotter without manually recording the data.

4.3 Direct Digital Control Program Package

4.3.1 Direct Digital Control

Direct Digital Control is an advanced form of

computer process control. Computer generated signals are used to control the process directly, without the intervention of analog controllers.

The most important characteristics of direct digital control can be summarized as follows.

Controller adjustments can be carried out with ease from the process operator's console. New controllers can be added to an existing system by simply inputting the appropriate parameters from punched cards. Measurements, setpoints, and controller outputs can be determined with high precision. An almost unrestricted range of controller constants is available. Advanced control algorithms can be applied through existing communication links between the digital control software and user written programs.

4.3.2 Direct Digital Control Program

The direct digital control program is a system program that resides permanently in core storage. It monitors all data acquisition and control functions of the computer. The initial activation of the program proceeds via the main console and the card reader. Its execution is initiated by interrupts at intervals specified by the user. In this particular installation, it was executed every second. Special communication programs are available that service the process operator's console and print alarm messages. Communication with the direct digital control program is provided

for by utility programs. With the aid of these utility programs, the user can load the process variable table or parts thereof from input devices into core storage, and obtain the contents of the table through output devices at any time. The table will be described in Section 4.3.3.

A typical set of functions handled by the direct digital control program are the following.

Measurements of process variables are obtained at user specified intervals, compared to given alarm limit values, filtered according to user requests, and stored in the process variable table. Valve commands are calculated using control algorithms chosen by the user. Commands are transmitted to the process via analog output equipment. Calculated controller outputs are stored in the process variable table for use in other variable records or programs.

4.3.3 Process Variable Table

The direct digital control program is organized around a single table of data referred to as process variable table. This table contains a variable record, also referred to as loop record, for each variable that is controlled, manipulated, or recorded. A variable record contains coded instructions, parameters, and logical flags that define the action to be taken by the direct digital control program with respect to the variable in question. The process variable table can be brought into core storage via the

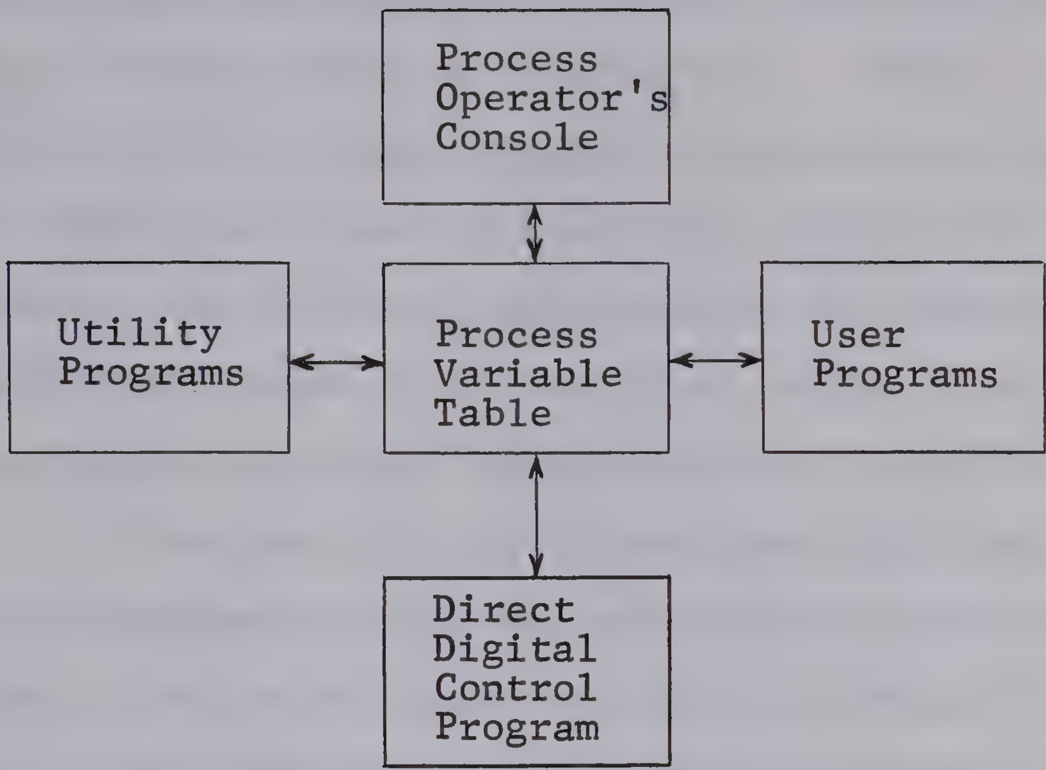
card reader in a prescribed format. Most entries in the table may be altered at any time from the process operator's console.

An example of the process variable table has been included in Appendix D. At execution time, the direct digital control program scans the process variable table to determine the action to be taken with respect to every single process variable at this particular instant. The parameters in the table will define the action required. In particular, they will specify the frequency at which each variable requires the program's attention. In addition, the table contains all information required by the program for servicing each process variable, such as the source of data, the range of measurements to be expected, alarm limits, control parameters, setpoints, filter constants, and the destination of the output.

Although the process variable table is loaded into core storage and maintained by utility programs, it can be altered at the discretion of the user from the process operator's console. A special set of sub-routines is provided that allows user written programs to access the process variable table, to extract information from it, or to supply information to it. All parameters contained in the table may be displayed on the console in hexadecimal or decimal notation, or in engineering units. The various means of communication with the table are illustrated on Figure 4.2.

Three basic kinds of variable records have been employed in the present study; namely, data acquisition records, control records, and data accumulation records. A data acquisition record references a variable that is to be measured only. At time intervals specified in the record, the program obtains a measurement and stores it in the record from where it is available to other records or user programs. A variable that is to be controlled or manipulated is represented in the table by a control record. This record contains specifications for the required control action and for the destination of the calculated output. The output may go directly to a final control element, or it may be placed into the record of another variable. Data accumulation records may be used to collect a series of measurement values from any process variable. Certain options may be exercised as to the destination of the accumulated data when the record is full. The data may be disregarded or written into a file on disk, or the record may be taken out of service, thus saving the data it contains.

The calculation of control output values is based on error signals. If a variable is referenced by a control record in the process variable table, the program will perform control calculations as specified in the record. The basic control modes that may be selected are proportional, integral, derivative, or a combination thereof. The algorithms the program uses are listed in Appendix C. In addi-



Communication with the Process Variable Table

Figure 4.2

tion, a number of special routines are available that permit averaging of several variables, ratio control, and material integration. An output bias term is provided that permits results of advanced control calculations to be added to the controller output calculated by the standard routines. This feature has been used in the implementation of feedforward control.

4.3.4 Distinctive Features

Two characteristics of direct digital control that require specifications not encountered with analog controllers are the availability of digital filters and the discrete sampling techniques. In the present study, filters were

employed for two purposes. Firstly, they smoothed measurements to reduce noise or fluctuations. Filter constants were selected based on consideration of process time constants and then refined empirically. Secondly, filters were used in conjunction with control calculations to obtain a specific predetermined shape of a controller output curve. This latter application will be expanded on in Section 5.2.

The two kinds of filters that the direct digital control program provides are exponential filters and Union filters, the latter named after their inventor. The exponential filter reduces a step change in the raw measurement to an exponential curve in the filtered measurement, thus acting as a damper.

The Union filter uses a velocity limiting procedure. It allows the rate of change of the measurement to increase as the error increases. At steady state, it prevents the measurement received by the computer from passing a specified limiting distance from the setpoint in any one scan period. In this way, spikes in measurements are effectively reduced. This filter was used in the nested control configuration of the feed system, where it prevented noise from propagating.

The algorithms for the filter actions are listed in Appendix C.

It is expected that digital filters will become increasingly important in future work, as more practical

experience will more closely define their assets and their range of applications.

As has been mentioned, the frequency at which variable records are polled by the direct digital control program is specified by the user. In the present study, these intervals were determined according to the expected rate of change of different variables, and according to the destination of the measurements. They ranged from one second for flow controllers up to 64 seconds for data accumulation records. In systems of nested control loops, the inner loops were polled more frequently than the outer loops.

The program was basically capable of reading up to 100 analog inputs per second. Since the computer was simultaneously serving other process users who required analog input readings, it became desirable to distribute the work load for the evaporator evenly over long periods of time. The polling sequence resulting from these considerations is shown in Appendix D.

5 DIGITAL CONTROL OF PRODUCT CONCENTRATION

5.1 Mathematical Model

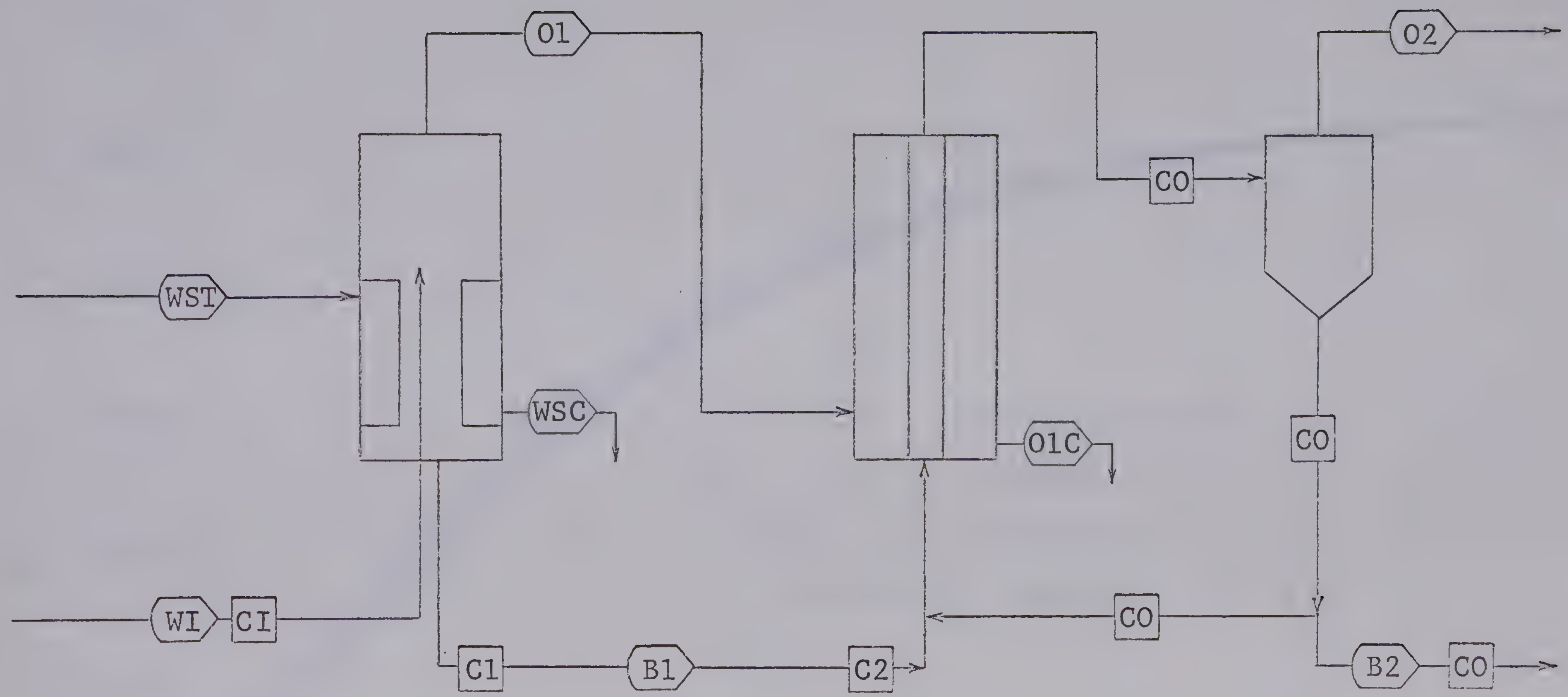
A mathematical model was needed for both the feed-forward and the inferential controller that were to be incorporated in the digital control system. The model was intentionally kept simple, in order not to obscure the concepts of implementation. Material balances and experimental steam economy formed the basis for this model, which was developed with reference to Figure 5.1. In this model, it was assumed that the feed entered the first effect at a constant temperature, and that the levels in both effects were constant at all times. The former assumption held true throughout the experimental program, as feed temperature controls were available. The latter assumption did not hold true during the experimental period, as with the successful implementation of averaging level control, both levels were allowed to vary. As a consequence, it was expected that the model would give a slightly faster response than the process. The dynamics of heat transfer in both effects was assumed to be negligible compared to the other process time constants.

Overall material balance

$$WI = O1 + O2 + B2 \quad (\text{since } WST = WSC \text{ and } O1 = O1C) \tag{5-1}$$

First effect solute balance

$$WI*CI = B1*C1 + H1*\frac{dC1}{dt} \tag{5-2}$$



Flows

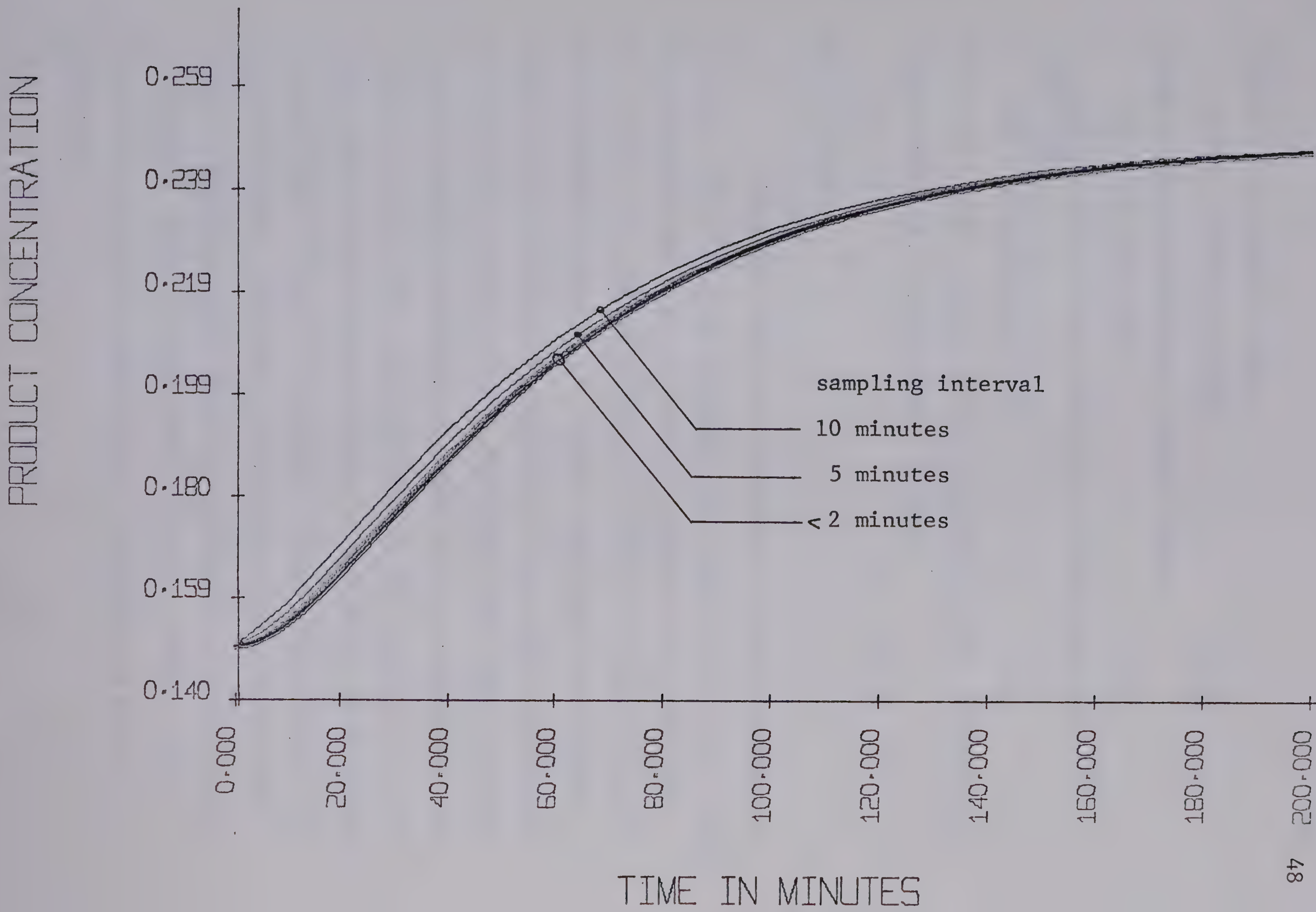
Concentrations

H1 - Hold-up in First Effect

H2 - Hold-up in Second Effect

STEC - Steam Economy

Figure 5.1 Basis for Process Model


PRODUCT CONCENTRATION
0·259
0·239
0·219
0·199
0·180
0·159
0·140
0·000
20·000
40·000
60·000
80·000
100·000
120·000
140·000
160·000
180·000
200·000
TIME IN MINUTES
sampling interval
10 minutes
5 minutes
< 2 minutes

Second effect solute balance

$$B1*C2 = B2*CO + H2*\frac{dCO}{dt} \tag{5-3}$$

Steam economy based on an overall balance

$$O1 + O2 = WST*STEC \tag{5-4}$$

For equally distributed evaporation (experimentally verified)

$$O1 = 0.5*WST*STEC = O2 \tag{5-5}$$

The 0.86 minute transportation lag between the effects has been neglected in this model, so that

$$C2 = C1 \tag{5-6}$$

at all times.

Once the model was developed, it was necessary to select a time interval at which the output of the model would be evaluated and used for control purposes. By simulating the process and plotting response curves obtained for various computation intervals, it was shown that the response curve does not change if intervals of less than one minute are used. The results of the simulation are shown on Figure 5.2 for sampling intervals ranging from ten minutes to less than one minute. The curves represent the response to a 20% step change in feed concentration. Considering these results, it was decided to use a one-minute computation interval for all simulations. The effect of sampling the measured variables at one-minute intervals is equivalent to introducing a pure time delay of approximately thirty seconds. This is the same order of magnitude as the neglected process time delay.

Equations (5-1) to (5-6) can be combined to yield two differential equations expressing the concentrations in both effects as functions of time.

$$\frac{dC2}{dt} = \left[\frac{1}{H1}*WI\right]*CI \quad - \quad \frac{1}{H1}*\left[WI - 0.5*STEC*WST\right]*C2 \qquad (5\text{-}7)$$

$$\frac{dCO}{dt} = \frac{1}{H2}*\left[WI - 0.5*STEC*WST\right]*C2$$

$$- \frac{1}{H2}*\left[WI - STEC*WST\right]*CO \qquad (5\text{-}8)$$

Equations (5-7) and (5-8) constitute the mathematical model of the process that was used to design both the feedforward and the inferential controller.

5.2 Feedforward Controller

5.2.1 General Description

It was desired to provide feedforward compensation for feed flow as well as concentration disturbances. The digital controller had to recognize these load disturbances and correct the steam flow to such a degree as to prevent them from propagating through the process. The entire feedforward controller was made up of three variable records that could be manipulated easily from the process operator's console. No additional programming was required. The controller is shown schematically on Figure 5.6.

5.2.2 Compensation for Feed Concentration Disturbances

The relation between the feed concentration and the steam flow is a first order lag system due to the liquid hold-up in the first effect. Equations (5-7) and (5-8) were used as a basis to derive this relationship. The mathematical development is presented in Appendix C. The resulting relation in terms of perturbation variables is as follows:

$$\frac{dWST'}{dt} = \left[\frac{\overline{WI}}{H1} * \frac{\overline{WI} - 0.5 * STEC * \overline{WST}}{0.5 * STEC * \overline{C2} - STEC * \overline{CO}} \right] * CI'$$

$$+ \left[\frac{STEC * \overline{CO}}{H1} * \frac{\overline{WI} - 0.5 * STEC * \overline{WST}}{0.5 * STEC * \overline{C2} - STEC * \overline{CO}} \right] * WST' \qquad (5\text{-}9)$$

It remained to assemble a digital controller that would produce the same output as a solution of this equation. At least two methods were available to arrive at this controller. Firstly, the equation could have been programmed and solved on the computer, either analytically, or numerically, and the result transferred to the control record concerned. In the analytical case, the solution is available explicitly in general terms. Appropriate constants are substituted by the program at regular intervals, and the result is calculated. In the numerical case, the differential equation is used as such in the computation. Successive approximations yield the value of the dependent variable at the end of a given interval.

As for the second method, the direct digital con-

trol program contained provision for exponential filtering of a process signal. This feature made it unnecessary to resort to the use of a separate, special program in the present case. For large time constants, the exponential filter approximates a first order system, and the filter constant is a function of the process time constant and the sampling interval. The mathematical definition of the exponential filter is given in Appendix C, Table C-1.

To demonstrate the equivalence of the two methods, a test run was performed during which a step change in feed concentration was simulated and the filter output recorded. The output obtained from the filter as a function of time agreed clearly with the solution of the given differential equation. Figure 5.3 shows the comparison of the two methods. After this test, the consideration of FORTRAN programming was abandoned, and the feedforward controller was implemented as shown on Figure 5.6.

As the process model was linearized to arrive at the equation for the feedforward controller, the accuracy of the compensation was expected to diminish as the process moved away from the initially defined steady state. The method described provides for very simple redefinition of the steady state for the controller. If a new steady state value of the feed concentration is entered into the variable record from the process operator's console, a new reference state is defined for the controller. Control action will

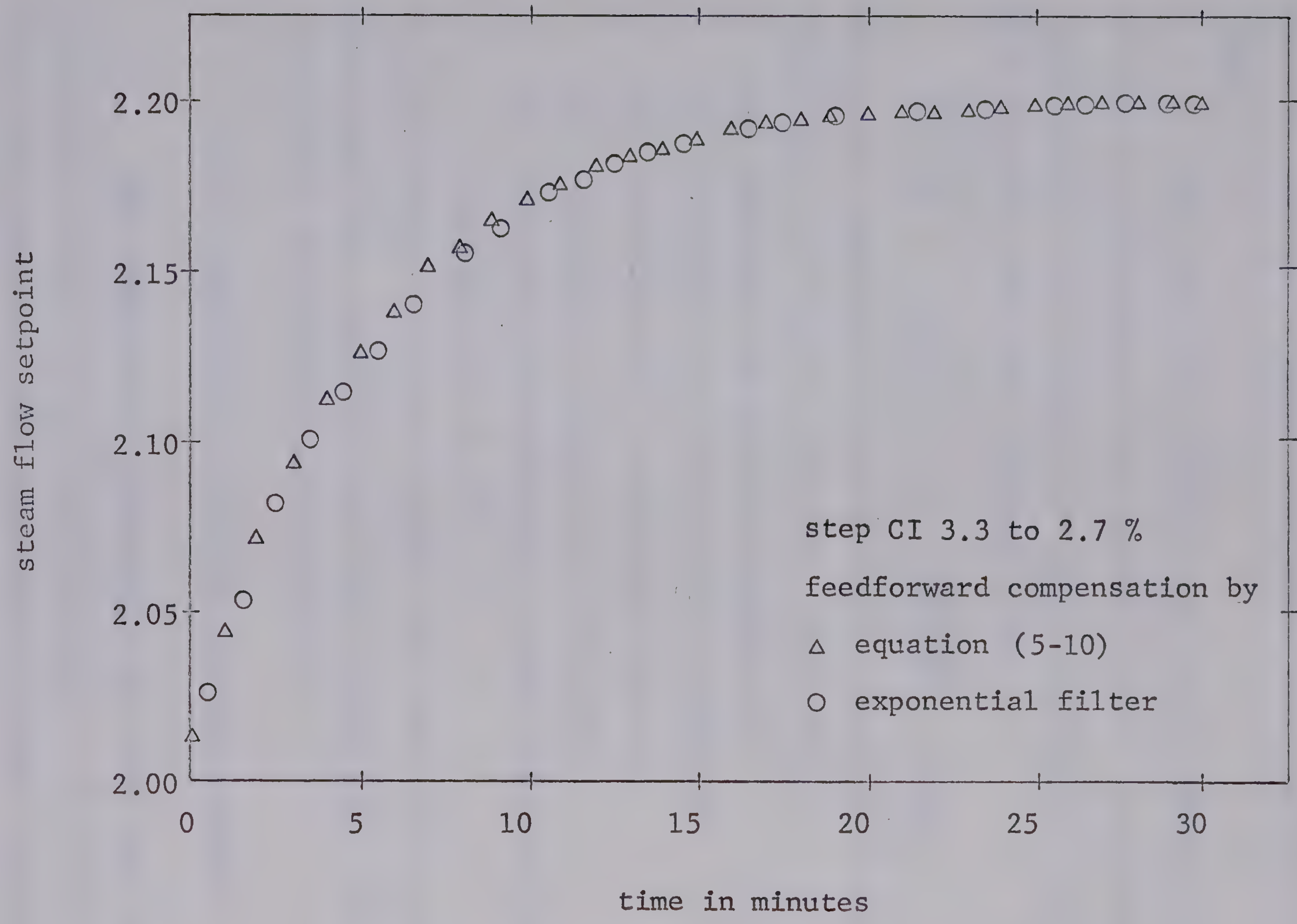


Figure 5.3 Two Alternatives for Feedforward Compensation

begin as soon as the measurement moves away from this latest value.

The pertinent constants for the feedforward controller were determined from equation (5-9). After Laplace transformation and substitution of the appropriate constants, this equation takes the following form:

$$\frac{WST\ (s)}{CI\ (s)} = \frac{K}{\tau s + 1} = \frac{-0.314}{5.3\ s+1} \qquad (5\text{-}10)$$

The dynamic compensation represented by equation (5-10) was implemented by using one DDC loop. This loop

(a) measured the feed concentration CI

(b) generated the perturbation variable CI'

(c) multiplied CI' by the gain factor K

(d) dynamically compensated (filtered)

(e) generated the deviation in steam flow WST' needed to compensate for CI'

The perturbation in steam flow WST' was added to the steady state value in a separate steam flow control loop. The implementation is described in Appendix D.

5.2.3 Compensation for Feed Flow Disturbances

The feedforward compensation for feed flow disturbances was determined from considerations of the level control scheme employed. The relationship between the feed flow and the steam flow will depend on the type of level control used in the first effect. If a tight level control is used, then the bottoms flow will respond immediately to a disturb-

ance in feed flow, and the steam flow will be related to the feed flow by a constant. In the case of averaging level control, the first effect is used as a damper, and the response of the bottoms flow is exponential. This dampening effect introduces a first order lag between the feed flow and the steam flow that had to be simulated in the feedforward controller

Two choices were available for the implementation, as is shown on Figure 5.6. In the first scheme, the feedforward controller read the feed flow disturbance. First order dynamic compensation as well as a gain factor as defined by equation (5-11) was applied, such that the output from this controller would be an exponential curve. The second scheme, that based on the simplified model, would produce the same end result, was to base the feedforward compensation on the bottoms flow from the first effect. In this case, the controller had to be given a gain factor only, since the first order lag was inherent in the physical relationship of B1' to WI'. Both possibilities were experimentally applied.

As it had not been attempted to describe mathematically the influence of averaging level control on the propagation of load disturbances through the process, the exact relation between steam flow and feed flow was unknown. In order to determine appropriate gain factors for the feedforward controller, constant steam economy as calculated

from a material balance was assumed. Starting from an initial steady state, the gain factor was calculated on the basis that the fraction of feed evaporated in the process be the same at the initial and at the final steady state. The time constant introduced between steam flow and feed flow by the changing hold-up in the first effect was assumed to be the same as that existing between the feed concentration and the steam flow, i.e. approximately equal to the average hold-up time in the first effect.

These considerations led to the following transfer functions (10):

$$\frac{WST\ (s)}{WI\ (s)} = \frac{0.40}{5.3\ s + 1} \qquad (5\text{-}11)$$

$$\frac{WST\ (s)}{B1\ (s)} = 0.61 \qquad (5\text{-}12)$$

The constants were supplied to the controller in the same manner as in the case of feed concentration. As can be seen on Figure 5.6, the flexibility of the digital controller provided the choice of using either one of the two methods.

5.3 Inferential Controller

5.3.1 Process Simulation

Inferential control consists of controlling a process variable whose value is not measured, but inferred. In this study, it has been applied to the concentration of

the product flow from the evaporator. A mathematical model of the process was developed and programmed for the computer. System routines were available which permitted communication between user programs and the process variable table. For the simulation program, the process model given by equations (5-7) and (5-8) was linearized using perturbation techniques. The linearized version was then Laplace transformed such that the model in terms of perturbation variables could be written in the following form:

$$C2(s) = \frac{CST1C}{s - CST1B} * CI(s) + \frac{CST9W}{s - CST9C} * WI(s) + \frac{CST2W}{s - CST2A} * WST\,(s) \qquad (5\text{-}13)$$

$$CO\,(s) = \frac{CST4C}{s - CST4E} * C2\,(s) + \frac{CSTTW}{s - CSTTF} * WI\,(s) + \frac{CST5W}{s - CST5D} * WST\,(s) \qquad (5\text{-}14)$$

It was found convenient to represent this process model in block diagram form as shown on Figure 5.4. Each of the six boxes on this diagram represents one term of equation (5-13) or (5-14).

Equations (5-13) and (5-14) were solved analytically, term by term. A program was then written to evaluate C2 and CO by substituting appropriate values into the analytical solution of these equations.

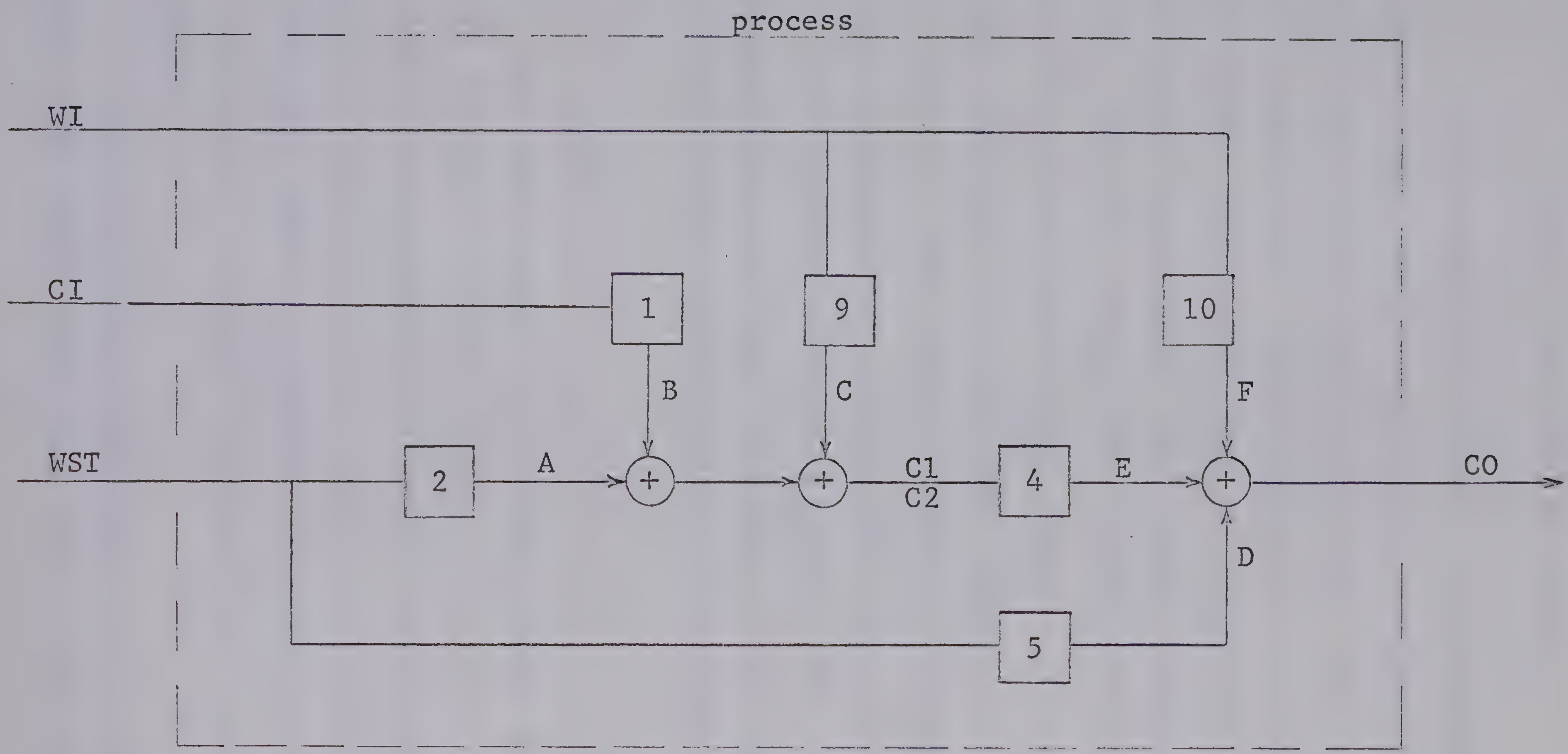


Figure 5.4 Process Model Block Diagram

5.3.2 Basic Implementation

The constants required for the linearized model were calculated using steady state values obtained from the process variable table at the start of each run. All programs are listed in Appendix E. The constants were calculated by the program DIANE, and the solutions of the equations were evaluated by the program COLOR. This latter program was called into core from disk storage twice during the specified control interval by a repetitive calling routine. It performed two different functions on an alternate basis. At the beginning and at the end of each control interval, the values of process input variables were obtained from the process variable table. At the midpoint of each interval, the product concentration was calculated from the model and its value transferred into the measurement of the concentration control loop.

Considering the control interval from time nT to (n+1)T, the measurement of the process inputs was done at time nT. At time $(n+\frac{1}{2})T$ the program calculated the value of the product concentration, CO((n+1)T) based on the inputs measured at time nT, which was sent to the DDC control program. This was done to spread the computing load over the full interval and partially compensate for the delay due to sampling.

In the analytical solution of the differential equations, the value of the time was constant and equal to

the interval between two successive calls to the program. Thus, the result obtained at every call represented an exact solution of the equations at this particular instant. All constants as well as the process vector consisting of inputs, outputs, and intermediate quantities on Figure 5.4, were written into a special section of core storage called the skeleton common area. The vector was updated at every call to the program, and the latest values used for calculations at the next calling instant.

As no program interrupt levels were available at the time the experimental work began, queuing routines were used to bring the program into core for execution. Thus, although the program was assigned the highest possible priority in the queue, it could not interrupt other process programs executing at the time of the call. The length of the delays caused was not known, but it was not expected that it frequently exceeded the time lapse between two successive calls.

5.3.3 Program Structure

The information flow within the inferential control system is shown on the logical flow diagram of Figure 5.5. The program consisted of a number of interconnected core loads. The functions performed by these core loads are summarized in Table 5.1. There are several features of this system that merit further discussion.

Firstly, the division into parts made it easy to

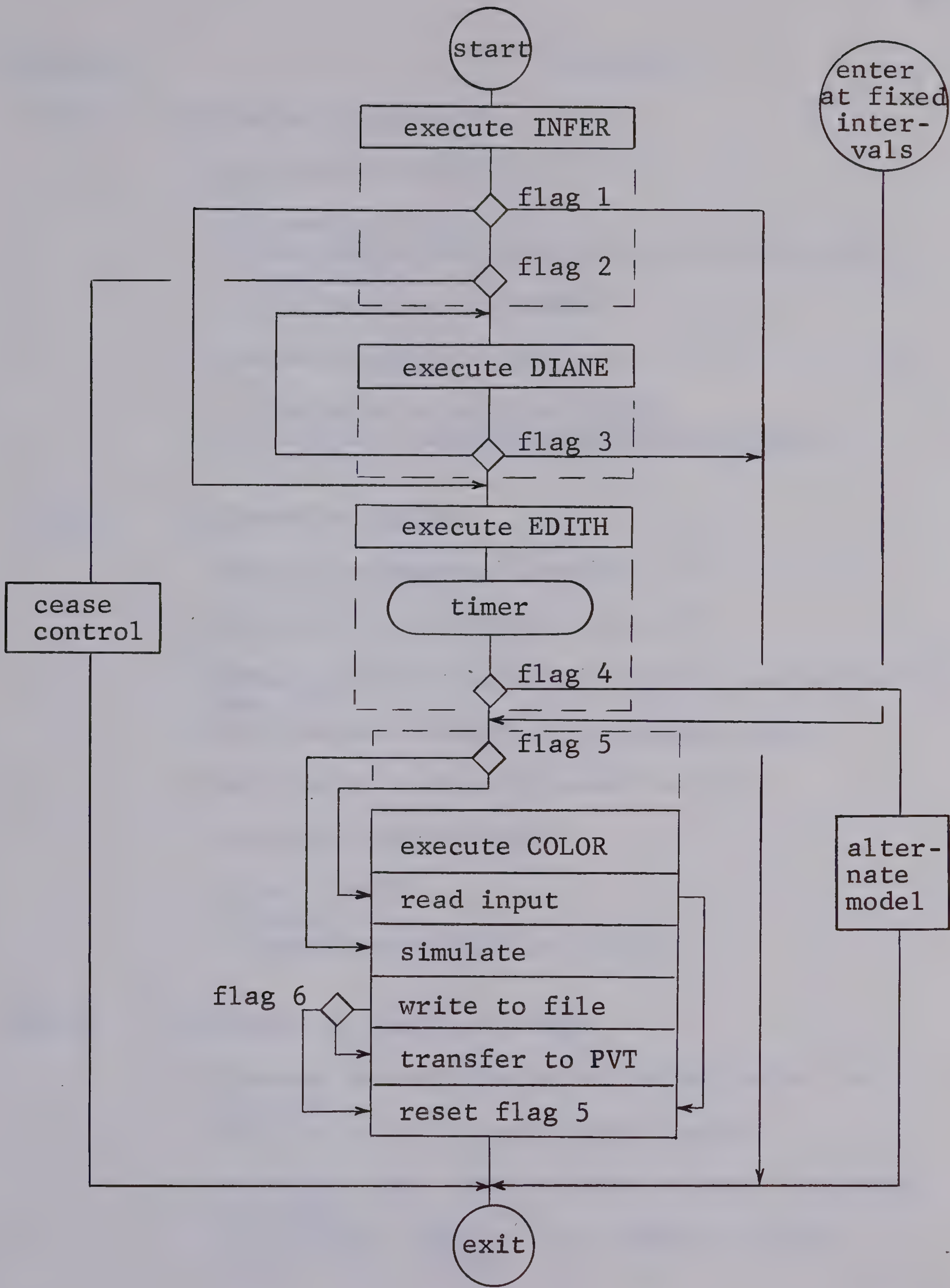


Figure 5.5 Information Flow Diagram

Programs	Functions
INFER	Initiated by operator 1. Operator enters flag 1 a) immediate exit b) calculate new constants for inferential controller c) use previous constants 2. Operator enters flag 2 a) continue by calling DIANE b) turn off timer and terminate program
DIANE	Initiated by INFER 1. Obtains parameters from POC 2. Obtains process inputs from PVT 3. Obtains present value of product concentration from PVT 4. Calculates constants for process model 5. Calculates initial process vector 6. Operator enters flag 3 a) call EDITH b) exit c) repeat execution of DIANE
EDITH	Initiated by DIANE or INFER 1. Operator enters flag 4 to select model to be used 2. Flag 5 is set to read input values 3. Requests time delay between process and model from POC 4. Initializes timer to the desired calling interval for COLOR

Table 5.1 Program Summary

Programs	Functions
COLOR	Initiated by timer Timer is automatically reset by the timer routine 1. Obtains present values of process inputs or (according to flag 5) Simulates process to obtain product concentration 2. Writes value of product concentration into a disk file or (according to flags 5 and 6) Transfers value of product concentration to feedback controller 3. Resets flag 5

Table 5.1 continued

modify the program. Any section could be removed and replaced by a different version without affecting the other sections.

Secondly, provision was made for the use of different process models. An alternate model could be called simply by changing a logical flag number. It is expected that this feature will be made use of in future work with this equipment.

Thirdly, multiple entry points to the timer called program COLOR were provided. This feature allowed for different functions to be performed at different calls from the timer.

Fourthly, the most important feature was the possibility of repetitive execution at fixed intervals. Once the timer was initialized, it queued the simulation program at specified intervals until it was turned off.

Fifthly, the program COLOR would write the results of the simulation into a file on disk from where they would be transferred as product concentration measurements to the feedback controller. Obviously, the value so transferred did not have to be the last value in the file, but could be specified to be any value from the file history. It was therefore possible to base control action on past information obtained from the model. This feature was applied in experiments INFI2 and INFII2. The model had been found to respond to disturbances seven minutes faster than the process. In an attempt to cancel this lead, the controller was supplied

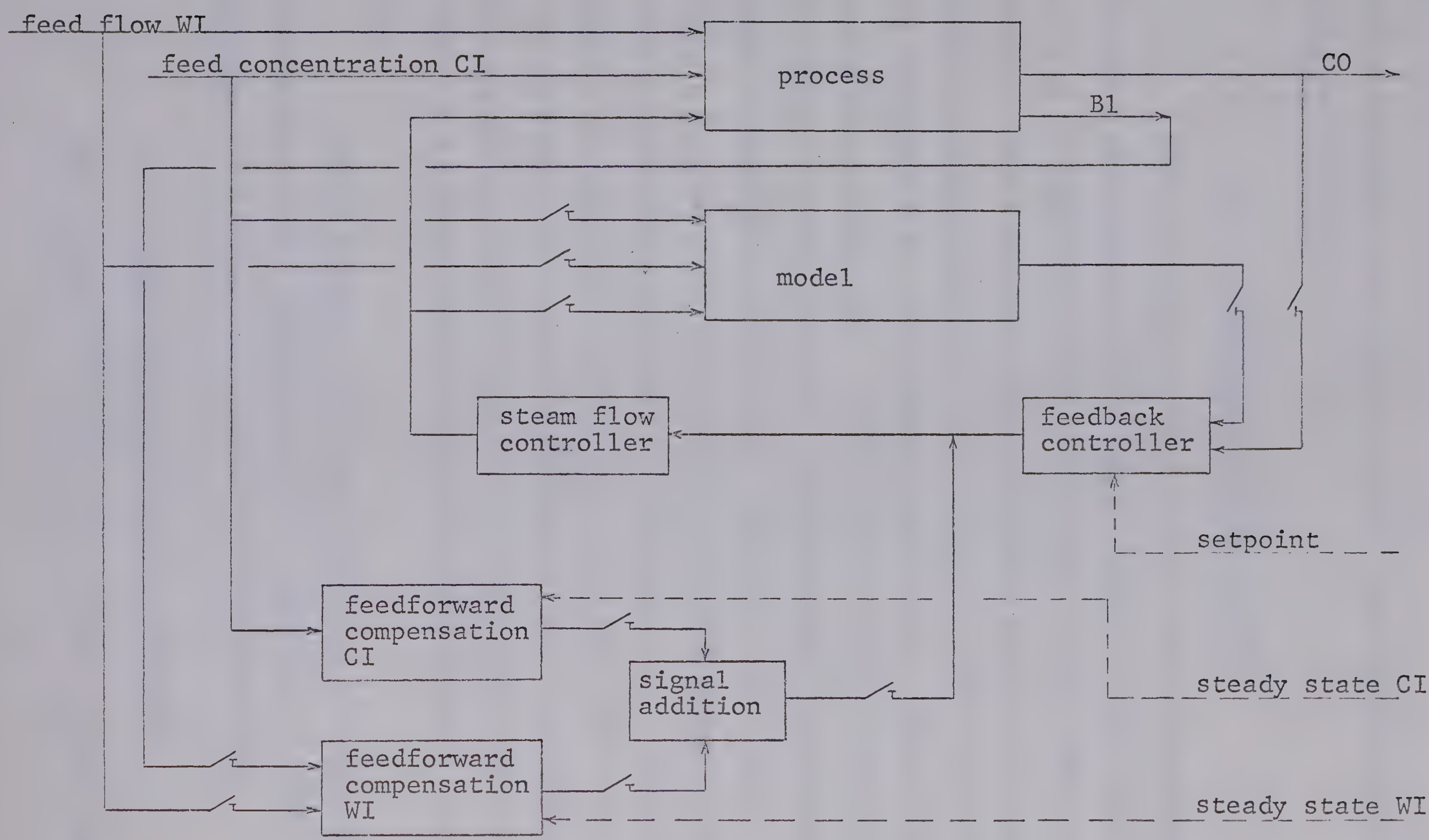


Figure 5.6 Product Concentration Control System

with measurements seven minutes after they had been calculated by the model.

5.4 Options for Control Action

The controller shown on Figure 5.6 was structured in such a way that any one component could be used individually or in combination with one or more of the other components. The inferential control system essentially substituted a mathematical model for the process that would supply the measurement for the feedback controller. Either the real process measurement or the inferred measurement could be used for control purposes, and feedforward compensation could be applied in either case. Moreover, the inferential control program shown on Figure 5.5 could be executed without actually performing control. In this case, the calculated product concentration was written into a disk file for plotting purposes, but was not transferred to the controller. This scheme permitted comparisons between model and process behaviour at times when the model was not involved in control.

These features made the controller an extremely versatile and well functioning unit. Reference is made to Appendix D for details of controller implementation.

5.5 Direct Digital Control Arithmetic

The operation of the direct digital control pro-

gram and the nature of variable records in this system are such that arithmetic and numerical operations may be performed by judiciously interconnecting a number of variable records. The method of using an exponential filter algorithm in the implementation of feedforward control was one example of this use of direct digital control operations. In general, "standard" variable records may be used to perform several operations, such as multiplication, division, addition, subtraction, or numerical integration.

The combination of actual control action with this type of manipulation is an important asset of direct digital control systems. It permits the implementation of feedforward and other advanced control actions without additional programming. The efficiency of this method of manipulation depends on the type of application and on the number of variable records involved.

The following is an illustration of several possible manipulations:

1. Proportional control algorithm

   output = Kp * [meas. - setpt.] + bias

   a) subtraction: bias = 0, Kp = 1

   b) addition: Kp = 1, setpt = 0

   c) multiplication: setpt = 0, bias = 0

2. Proportional - integral control algorithm

   output = Kp * [meas. - setpt.] +

   Kp * KI * [meas. - setpt] + REST

where REST is equal to the sum of Kp * KI * [meas - setpt] evaluated at all previous control intervals.

a) same manipulation as in 1.

b) integration: setpt = 0, KI = sampling interval delta t, Kp = 0 to eliminate the proportional term (a value of Kp = 1 is then automatically selected for the integral term)

$$\text{output} = \sum \text{meas} * \text{delta t}$$

3. Proportional - derivative control algorithm

$$\text{output} = \text{constant} * \left[\frac{d\ (\text{meas})}{dt}\ (\text{filtered}) + \text{proportional action as per 1.} \right]$$

filtered derivative action: setpt = 0, Kp = 1 or other values

$$\text{output} = \frac{d(\text{meas})}{dt}\ (\text{filtered})$$

4. Averaging control algorithm

$$\text{Output} = \sum \text{word (i) of loop (j)},$$ for all specified pairs of (i,j) values

5. Different poll and phase times can be used to perform iterative or sequential calculations. In the early stages of this study, the differential equation defining the action of the feedforward controller was solved by a group of six variable records that were arranged in such

a way as to perform a numerical integration.

6. Minor modifications to the data accumulation program would provide for implementation of pure time delays.

7. Model simulation

   A model of the form $\frac{K}{\tau s + 1} * \frac{1}{s}$

   could be implemented by two DDC loops. The first loop would perform the multiplication by the gain factor and the filtering action equivalent to the given first order system. The second loop would perform the integration.

8. Other feedforward models

   The simulation of transfer functions of the type $\frac{K * (\tau_1 s + 1)}{(\tau_2 s + 1) * (\tau_3 s + 1)}$

   is possible with the use of DDC loops. The efficiency of this method would have to be evaluated for a given application.

These examples serve to illustrate further the versatility of the direct digital control system and indicate that DDC programs can be used for other applications where the dynamic compensation or control scheme differ from those implemented in this study.

6 EXPERIMENTAL PROGRAM

6.1 Summary of Experiments

The complete experimental program is shown on Tables 6.1 and 6.2, and the complete results obtained from each experiment are given in Appendix A.

To ensure that all experiments were performed under similar process conditions, a steady state of reference was established. This reference state is defined in Appendix A. All load disturbances introduced into the process were 20% of the steady state value in magnitude, and used this steady state as a pivot.

The program was planned such that it would demonstrate the performance of all parts of the digital product concentration controller. These included the feedback controller responding to both measured and inferred signals, as well as the feedforward compensation for feed flow and concentration disturbances.

The feedforward compensation for flow disturbances was based on the feed flow signal in all experiments labelled FF II and FFB II, and on the first effect bottoms flow signal in experiments labelled FF IV.

Of the experiments using the inferential controller, runs INF I2 and INF II2 used a process model containing a time delay. For runs INF I and INF II, no time delay was implemented.

6.2 Experimental Procedure

All experiments listed on Tables 6.1 and 6.2 were conducted in three extended periods of operation. At the beginning of each of these periods, the process was brought to the same steady state. All disturbances were introduced from the process operator's console.

Data accumulation records were available to retain measurements of the most important process variables at specified intervals. These included the product concentration, the steam flow, the liquid levels in both effects, and the bottoms flows from both effects. From the accumulation records, data were collected in a disk file for later plotting. At the end of each run, all required diagrams were produced directly on the system plotter. The disk file was then reset for the next run.

At the beginning and at the end of each run, a material balance was performed around the major sections of the process, that determined the steam economy and the errors of closure. The results of this balance could only approximate the condition of the process at that particular instant, as the measurements used in the balance calculations were instantaneous values that could contain noise components. For the later part of the experimental period, a more sophisticated material and energy balance program developed elsewhere in this department was available. This program obtained several consecutive measurements of each variable

and used an average in the subsequent calculations. It also contained the option to adjust the process vector such that the error of closure disappeared.

The refractometer recording the product concentration was not reliable at all times. To produce a continuous chart output that was correct within the error of measurement, samples were taken manually at intervals of approximately seven minutes and analysed on a bench refractometer. Whenever the concentration obtained from this analysis differed from that shown on the recording chart, the instrument was adjusted on the spot so that it would indicate the correct value. The discrete steps visible on some of the response curves in Appendix A are a result of this adjusting procedure.

6.3 Supplementary Software

To permit the proper functioning of the procedure outlined in the previous section, several additional programs had to be written. They have been kept as general as possible and are expected to be used in future work with this equipment. At the end of each run, the program BEGIN was called to reset the disk file that collected data from accumulation records. The diagrams at the end of each run were produced using programs THIS and THAT. The material balances were performed by FDFWD. Listings of these programs are contained in Appendix E.

Disturbances / Controls Used	Feed Concentration		Feed Flow	
	step up	step down	step up	step down
feedback, measured signal	FBI2	FBI3	FBII2	FBII3
feedforward signal from feed flow (WI)	FFI2	FFI3	FFII2	FFII3
feedforward, signal from first effect bottoms flow (B1)			FFIV2	FFIV1
feedback, measured signal + feedforward (WI)	FFBI	FFBI2	FFBII3	FFBII2
feedback, inferred signal model with time delay		INFI2	INFII2	
feedback, inferred signal model without time delay		INFI	INFII	
open loop, no controls	OLI2		OLII2	

Table 6.1 Experimental Program

Experiment	Feed Concentration Weight Per Cent Triethylene Glycol	Feed Flow Rate lbs/min
FBI2	2.8 to 3.4	
FBI3	3.4 to 2.8	
FFI2	2.7 to 3.3	
FFI3	3.4 to 2.8	
FFBI	2.8 to 3.4	
FFBI2	3.3 to 2.7	
INFI	3.1 to 2.5	
INFI2	3.5 to 2.9	
OLI2	2.4 to 3.0	
FBII2		4.5 to 5.5
FBII3		5.5 to 4.5
FFII2		4.5 to 5.5
FFII3		5.5 to 4.5
FFIV1		5.5 to 4.5
FFIV2		4.5 to 5.5
FFBII2		5.5 to 4.5
FFBII3		4.5 to 5.5
INFII		4.5 to 5.5
INFII2		4.5 to 5.5
OLII2		5.0 to 5.5

Additional Experiments	Disturbances Applied
OLIV	WST 1.58 to 1.80
CONINF	1. CI 3.0 to 3.3 + WI 5.5 to 5.0
	2. CO (setpoint) 7.8 to 9.0
	3. CI 3.3 to 3.6
	4. WI 5.0 to 5.3
	5. WI 5.3 to 5.0

Table 6.2 List of Applied Disturbances

7 DISCUSSION OF RESULTS

7.1 Experimental Data

7.1.1 Scope of Experimentation

As has been stated at the outset of this report, it was assumed that the feedstream to the evaporator was not available for manipulation. It was the function of the control system to correct for disturbances in the feed stream by manipulating the steam flow.

The experimental program was planned to permit the evaluation of the following digital control systems:

basic feedback control

feedforward control

combined feedback and feedforward control

inferential control.

Both the flow rate and the concentration of the feed stream were disturbed by applying positive as well as negative step changes.

The results obtained have been summarized on Figures 7.1 to 7.4. For all experiments, the initial and final steady state operating conditions were maintained as close as practicable to the design conditions, and all step changes introduced were of the same magnitude. Therefore, direct comparisons of response curves are possible. All data obtained during the experimental program, including initial and final steady state conditions and transient res-

Response to a Step up in Feed Concentration

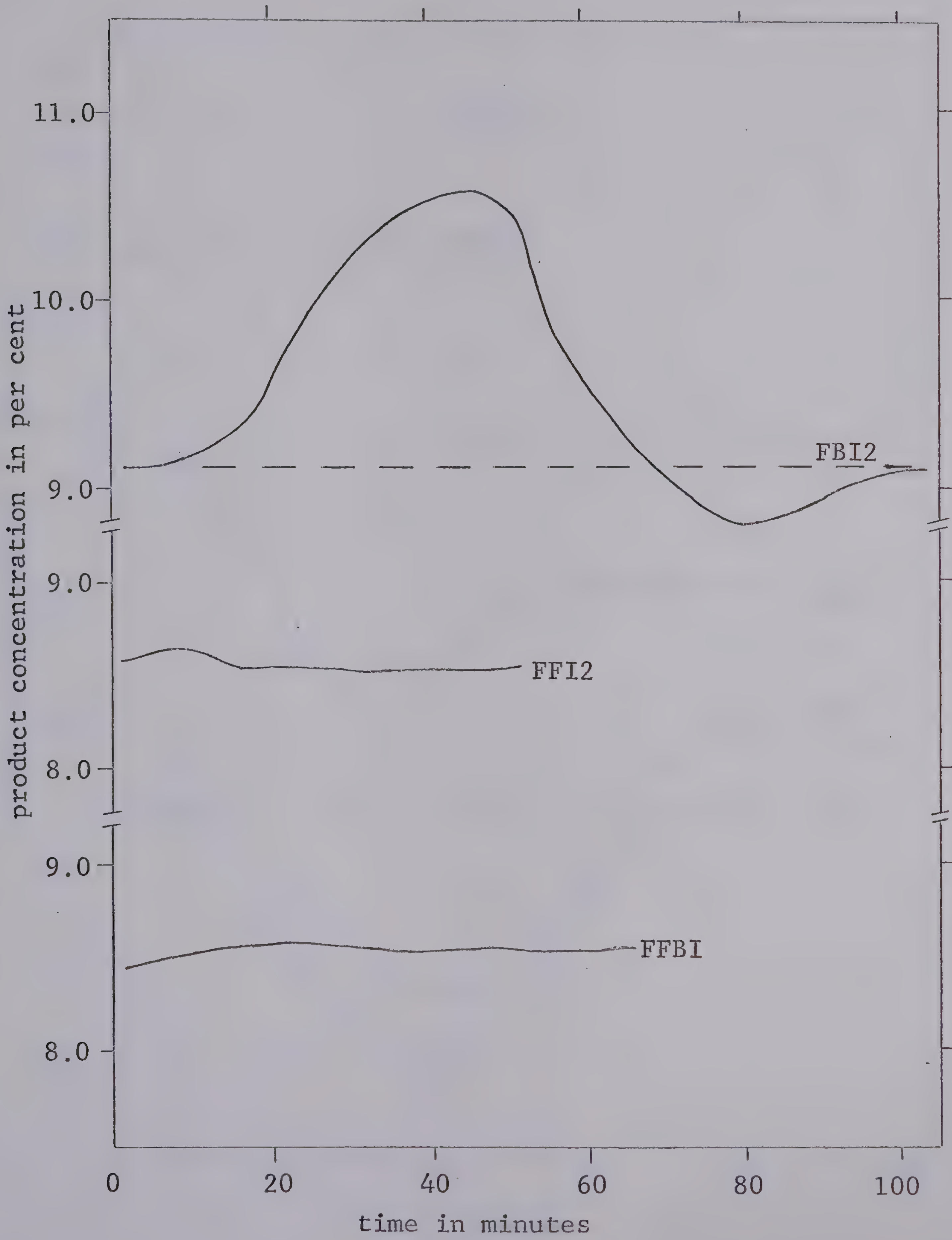


Figure 7.1

Response to a Step down in Feed Concentration

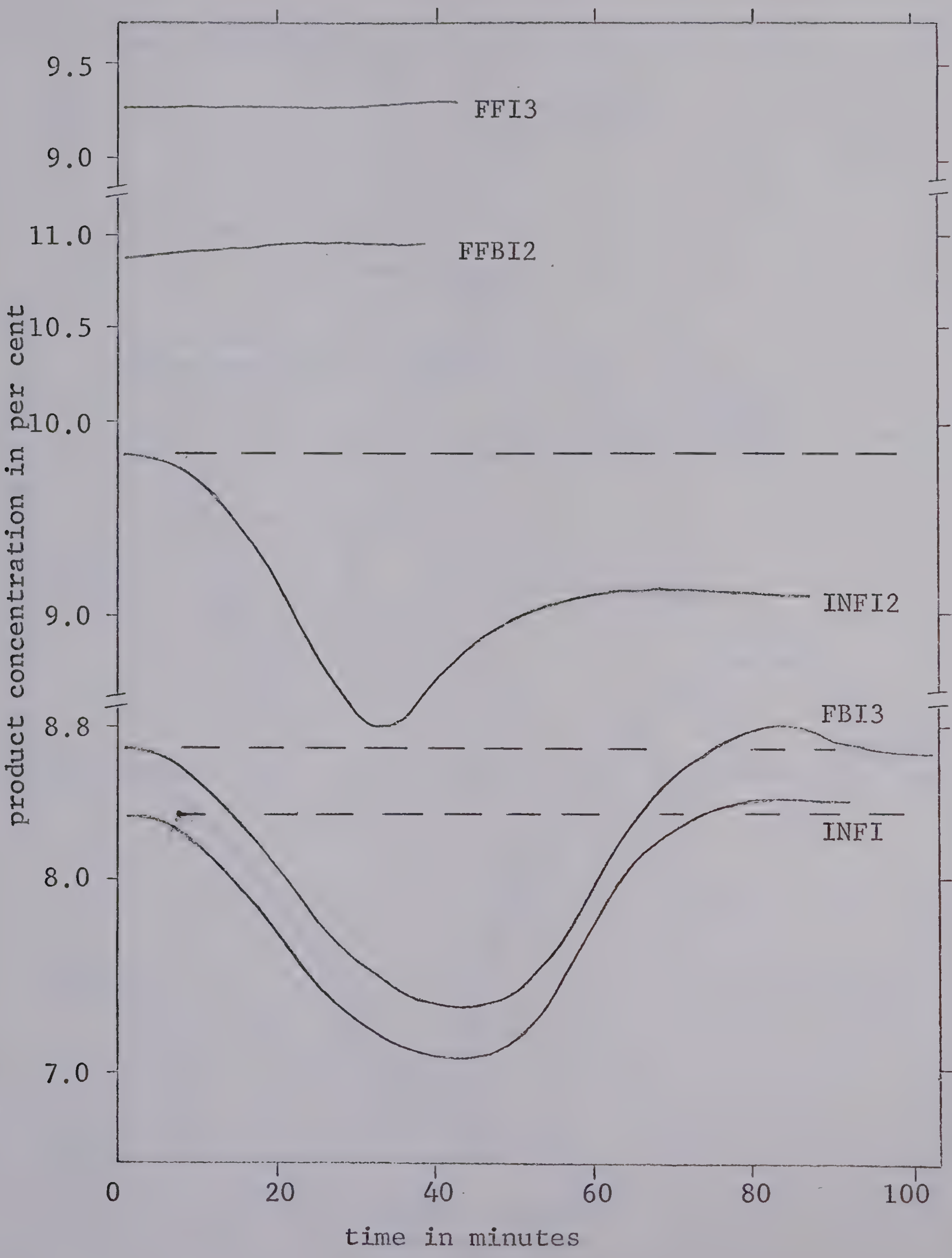


Figure 7.2

Response to a Step up in Feed Flow Rate

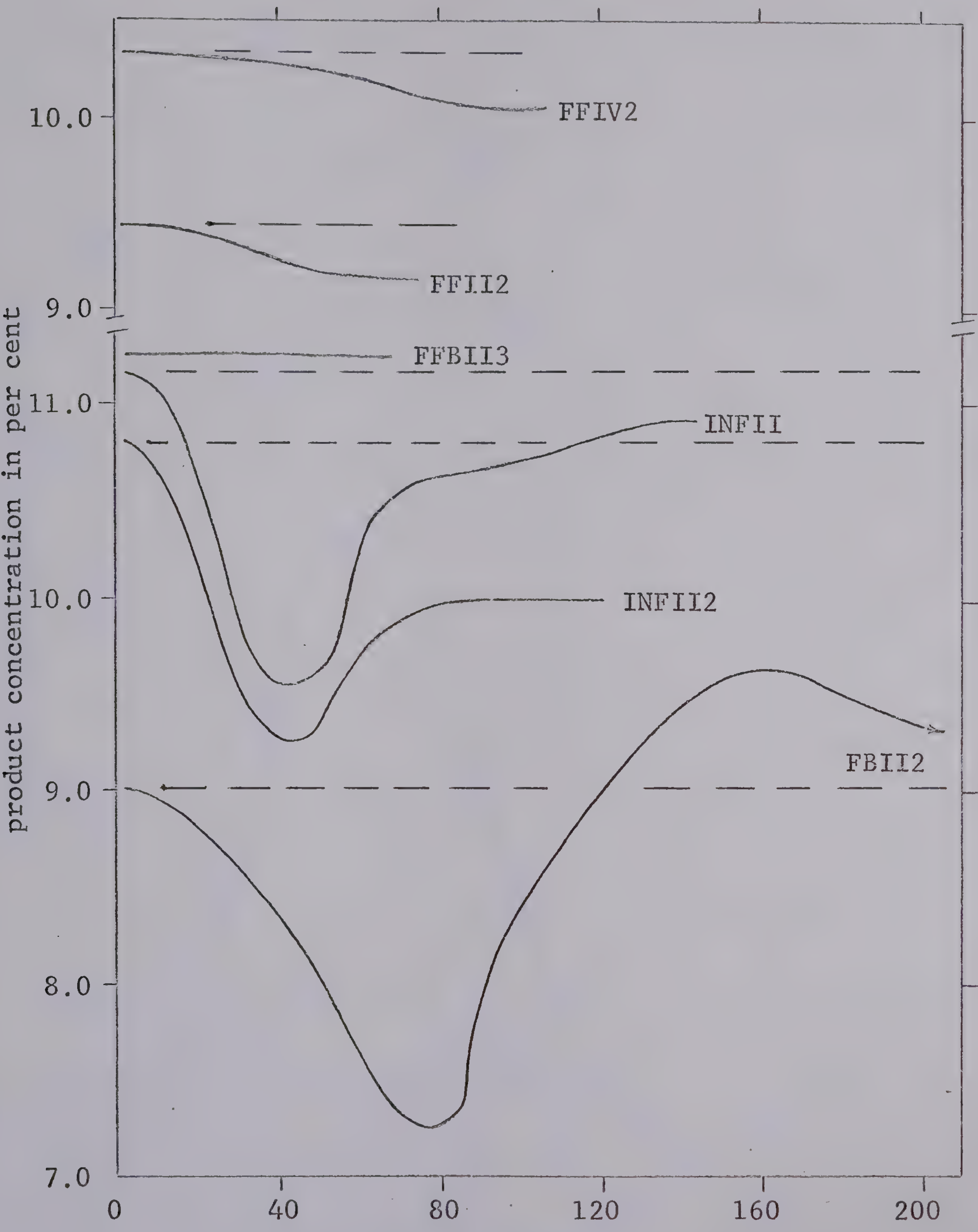


Figure 7.3

Response to a Step down in Feed Flow Rate

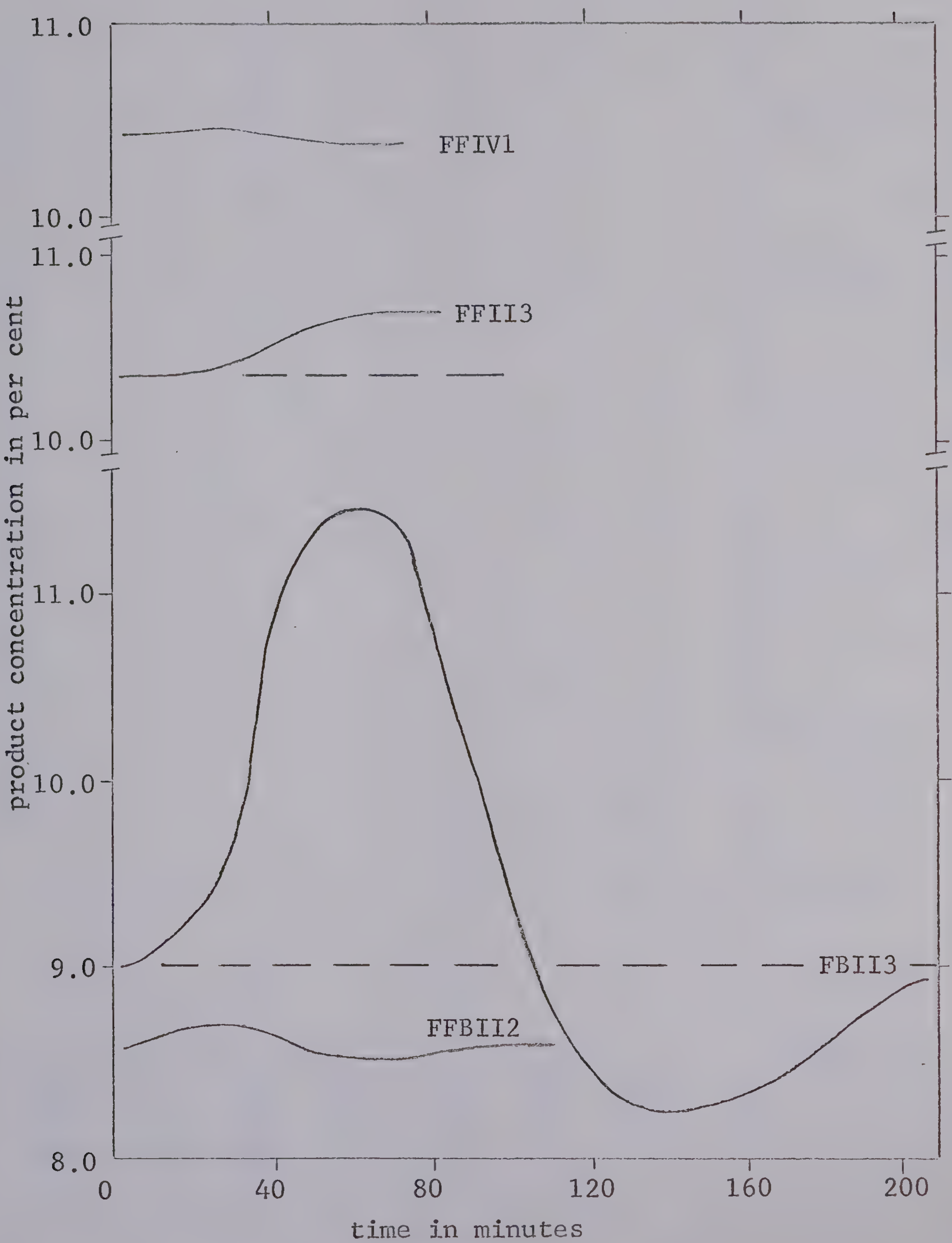
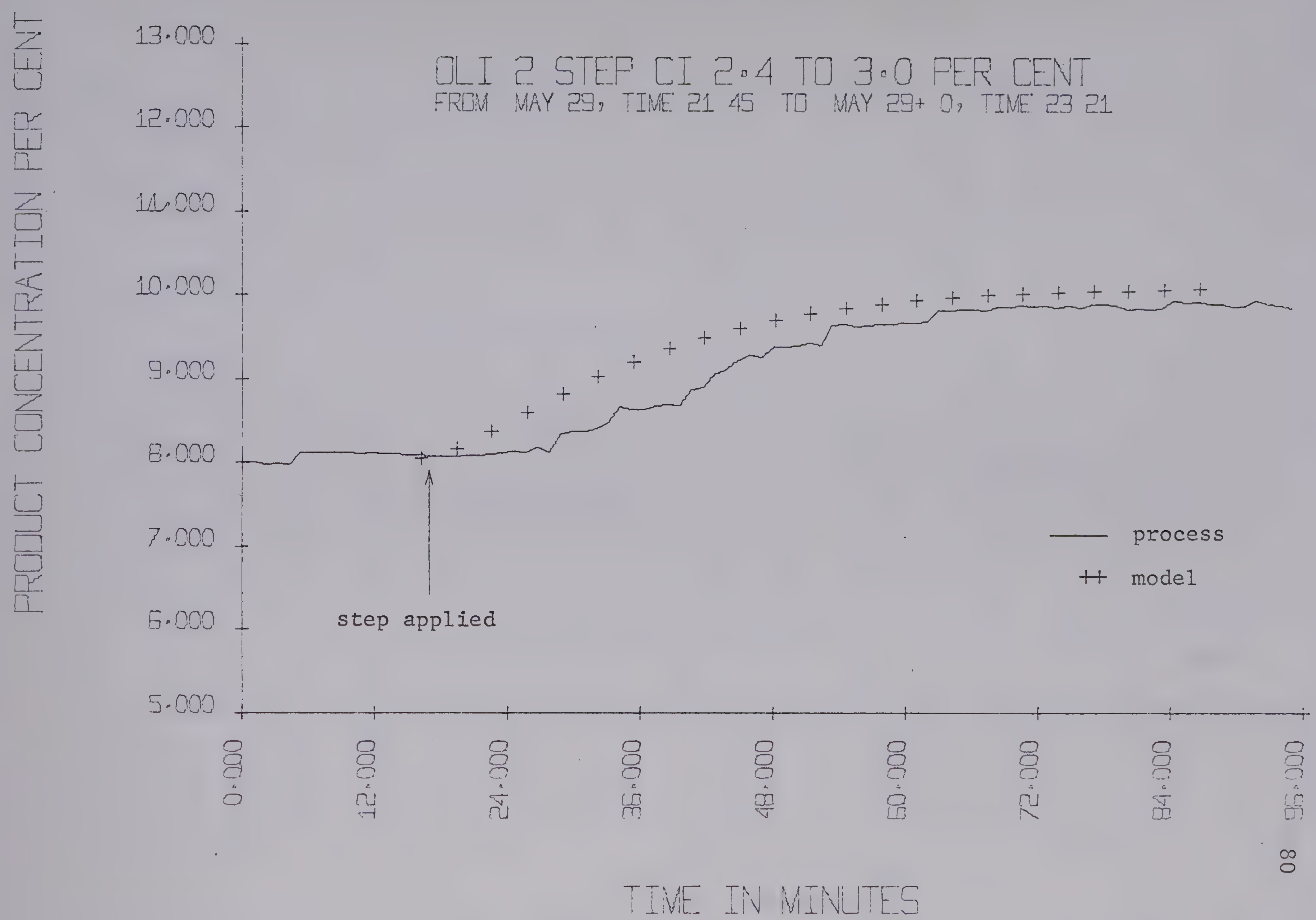
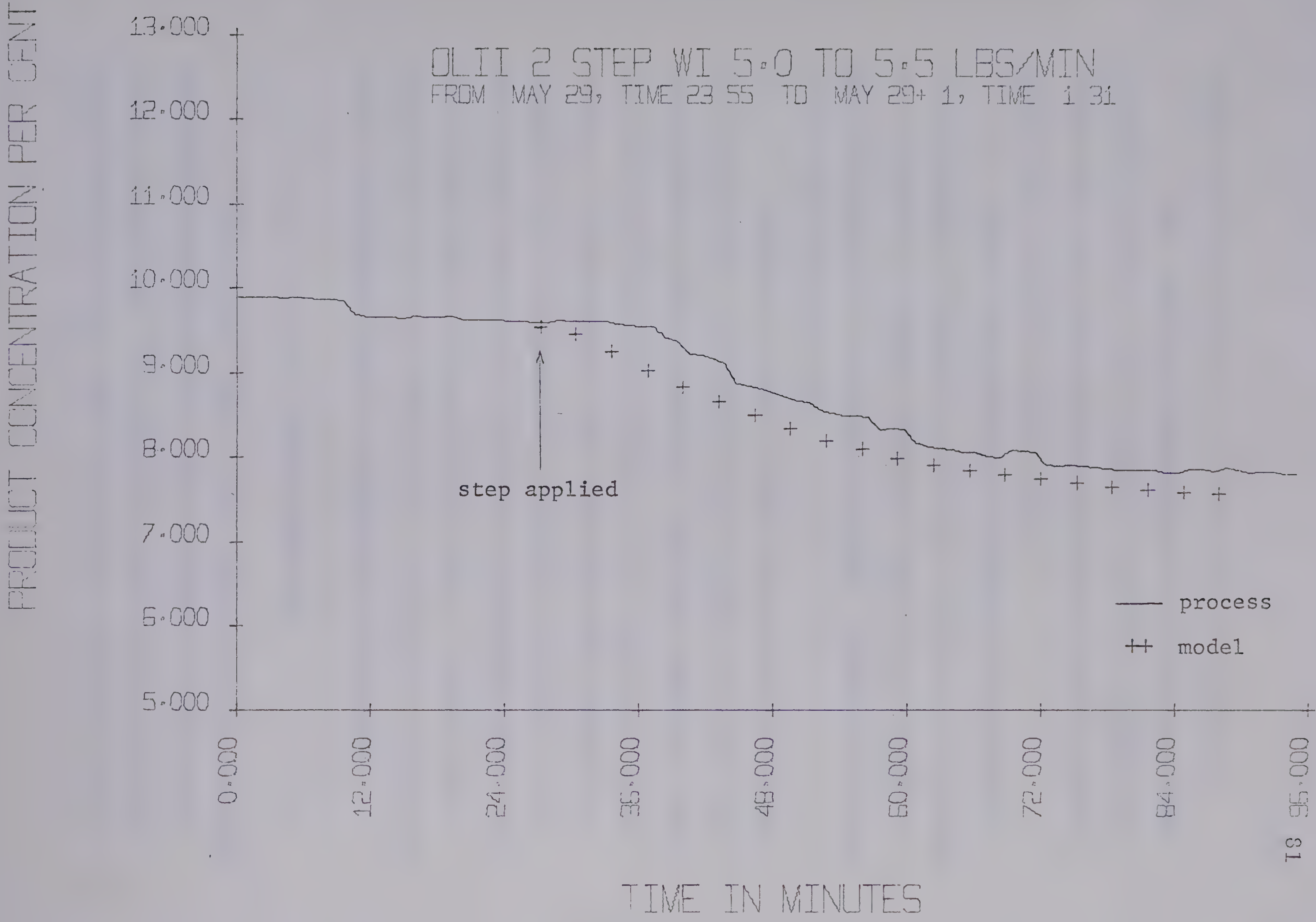


Figure 7.4


OLI 2 STEP CI 2·4 TO 3·0 PER CENT
FROM MAY 29, TIME 21 45 TO MAY 29+ 0, TIME 23 21
PRODUCT CONCENTRATION PER CENT
13·000
12·000
11·000
10·000
9·000
8·000
7·000
6·000
5·000
0·000
12·000
24·000
36·000
48·000
60·000
72·000
84·000
96·000
TIME IN MINUTES
step applied
process
model



OLII 2 STEP WI 5.0 TO 5.5 LBS/MIN
FROM MAY 29, TIME 23 55 TO MAY 29+ 1, TIME 1 31
PRODUCT CONCENTRATION PER CENT
13.000
12.000
11.000
10.000
9.000
8.000
7.000
6.000
5.000
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
TIME IN MINUTES
step applied
process
model

Figure 7.6

ponse curves for all runs, are included in Appendix A.

All runs were terminated as soon as a new steady state was reached. As is apparent from the diagrams, the feedforward runs required considerably less time to reach steady state than the feedback runs.

Direct comparisons of the present results with those obtained by Andre (2) and Wilson (31) are not possible, due to physical changes, different steady state conditions, and differences in disturbances applied. General agreement can be shown in most cases.

Figures 7.5 and 7.6 show the results of two open loop experiments. As had been anticipated, the model responded faster than the process. Andre had reported a similar situation which he attributed to the imperfectly mixed system. In the present case, two additional reasons are suggested for the discrepancy. Neither the delay in the feed piping nor the lag created by the averaging level control system were considered by the model. How well this simple model was suited for control purposes was determined by the experiments using the inferential controller.

7.1.2 Feedback Control

The general shape of the response curves obtained from the feedback runs (FB) shows evidence of the tuning criterion employed. The feedback controller was tuned for rapid damping and gave rise to analogous response curves

for all feedback runs.

The maximum deviation is larger for step changes resulting in a rising product concentration, as can be seen on Figures 7.1 and 7.4. This result is due to the varying sensitivity of product concentration to steam flow. As the concentration rises, as in runs FBI2 and FBII3, the change of concentration per pound of water evaporated increases. The difference in maximum deviation between runs FBII3 and FBII2 is much more pronounced than between runs FBI2 and FBI3, because the averaging level control system employed gave rise to varying liquid levels in the case of flow disturbances. A step down in feed flow causes the liquid hold-up to decrease, which entrails a larger change in concentration per pound of water removed.

The time required to regain steady state following a concentration disturbance was less than following a flow disturbance. This difference is due to the length of time taken by the liquid levels to return to their steady state values after a flow change.

The experimental results reported by Wilson show that in the case of a concentration change, cycling lasted for 180 minutes, and in the case of a flow change, steady state was not regained after 200 minutes. The higher throughput rates used in the present experimentation are considered a major factor in reducing the response time, as they decrease the hold-up time of the process. In addition, with the controller constants used here, cycling was not exper-

ienced. Andre did not report on concentration disturbances. In general, the shapes of his response curves for flow disturbances are similar to those obtained in the present work.

7.1.3 Feedforward and Feedback-Feedforward

The improvements in product quality control for this evaporator that can be achieved by the use of feedforward compensation had been indicated by Wilson. In the present work, it has been attempted to move toward general purpose software in the implementation of feedforward control. As the present experimental work confirmed the results obtained by Wilson, it is felt that as feedforward compensation is possible with standard software, it will become increasingly common in industrial processes.

The dramatic improvement due to this type of control can be seen from Figures 7.1 to 7.4. Concentration disturbances were compensated for completely by the feedforward controller. As no correction for off-set was required, the combined feedback-feedforward controller produced the same product quality as the feedforward controller. In the case of flow disturbances, the compensation was not complete, such that a slight off-set in product quality occurred that was eliminated by feedback correction for both positive and negative step changes.

Two methods of providing feedforward compensation for flow changes were applied. For runs FFII2 and FFII3,

the steam flow adjustment was determined by the relation

WST (s) = (K1 / (τs + 1))* WI (s)

For runs FF IV1 and FF IV2, the relation

WST (s) = K2 * B1 (s)

was used, where

WST = steam flow

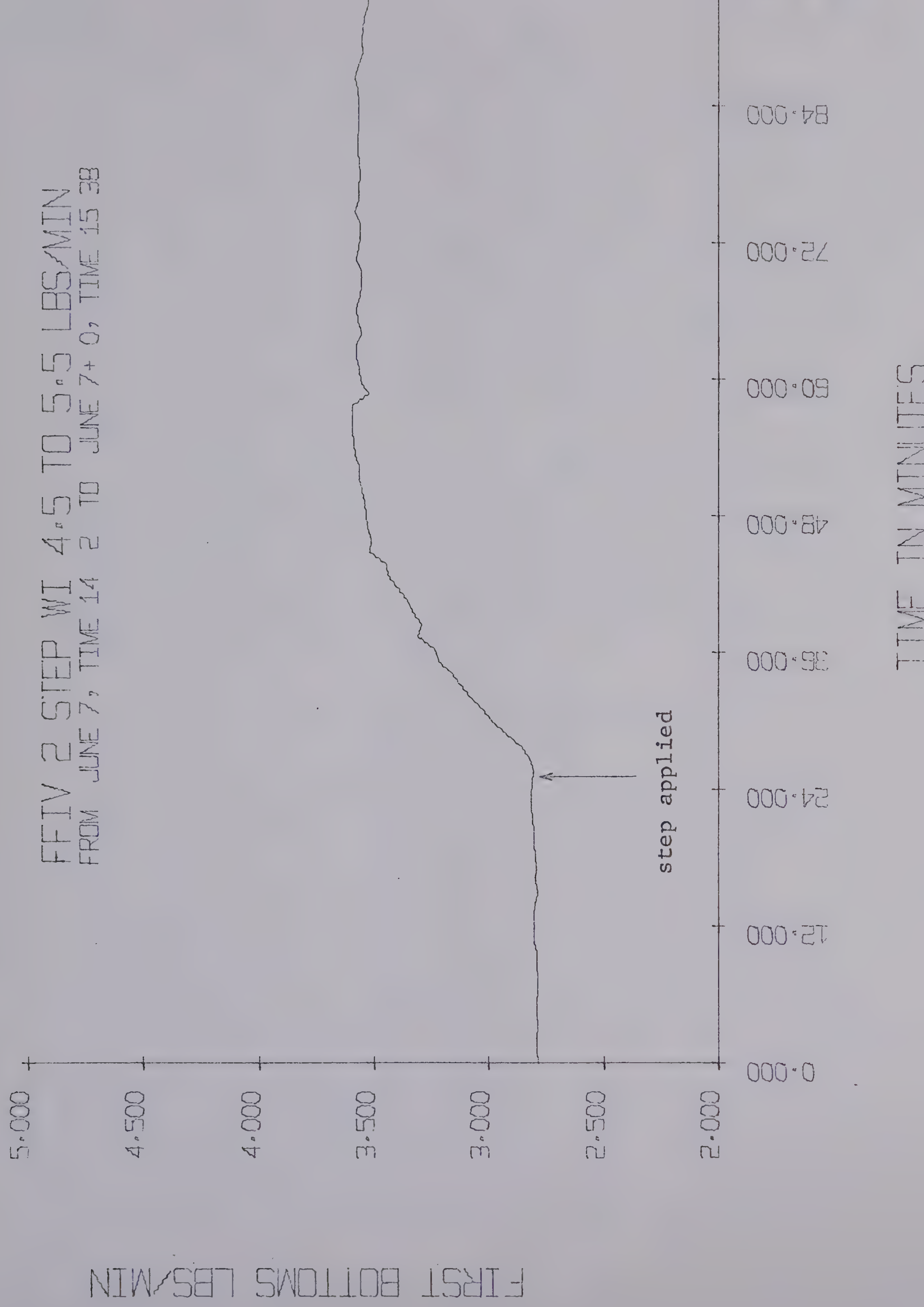
WI = feed flow

B1 = first effect bottoms flow

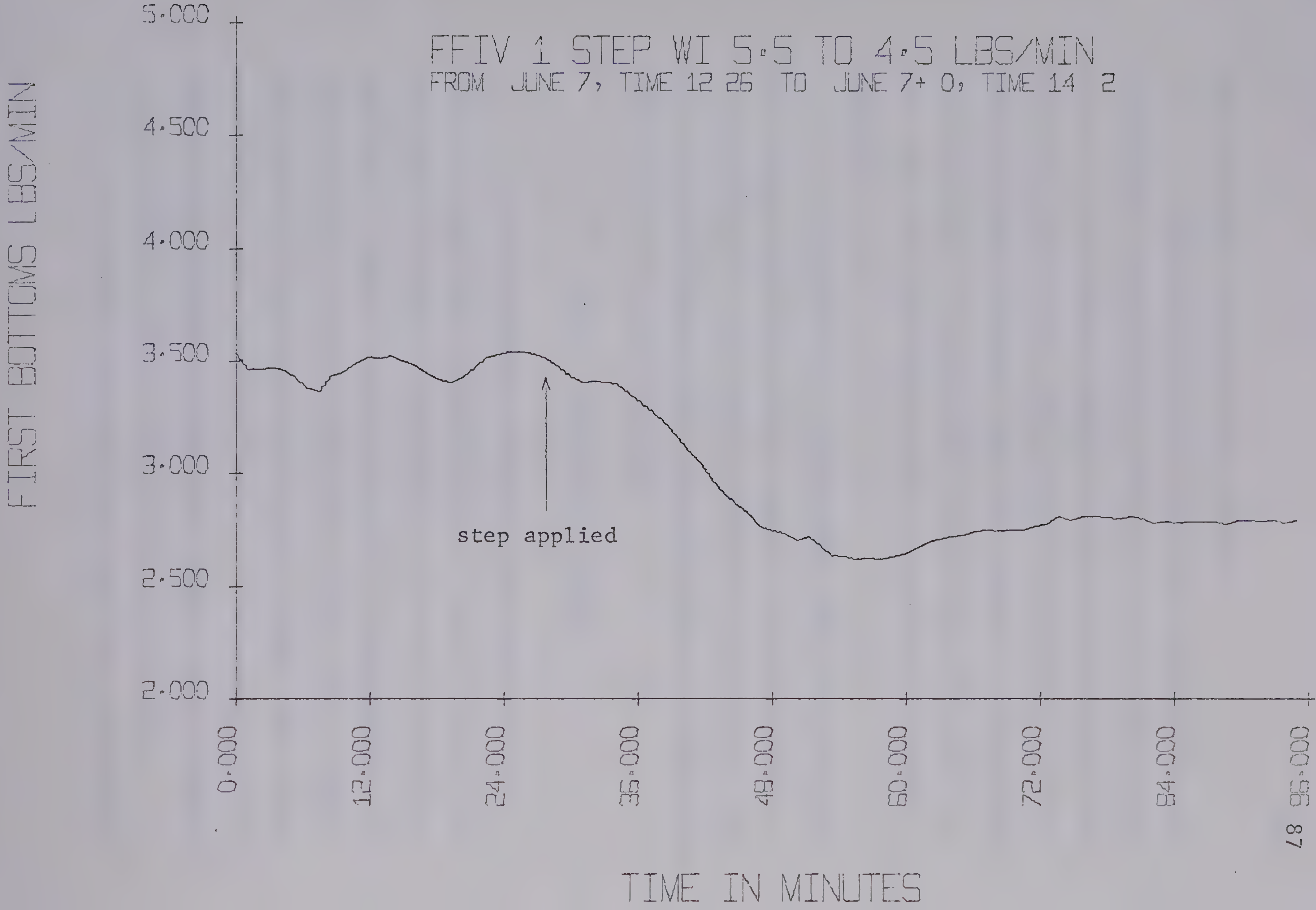
The response curves for runs FFII2 and FFII3 are analogous. The off-set occurring in these runs was due to an insufficient gain. If the constant K1, that had been found from theoretical considerations, were increased by an appropriate amount, the off-set would disappear.

As for runs FFIV1 and FFIV2 the steam flow was directly proportional to the first effect bottoms flow, Figures 7.7 and 7.8 have been included here for illustrative purposes. They show the behaviour of the bottoms flow for these two experiments. A careful examination of these figures plus recordings of the other variables did not provide any basis for the observed difference in response shown on Figures 7.3 and 7.4. Therefore, it must be conceded that the cause of the off-set occurring in experiment FFIV2 is unknown.

Small variations in process behaviour such as these suggest the continuing need for feedback correction in addition to feedforward compensation. This point is illustrated very clearly by experiments FFBII2 and FFBII3, in which feed-



FFIV 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 14 2 TO JUNE 7+ 0, TIME 15 38
step applied
FIRST BOTTOMS LBS/MIN
TIME IN MINUTES
5.000
4.500
4.000
3.500
3.000
2.500
2.000
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000


FFIV 1 STEP WI 5.5 TO 4.5 LBS/MIN
FROM JUNE 7, TIME 12 26 TO JUNE 7+ 0, TIME 14 2
FIRST BOTTOMS LBS/MIN
5.000
4.500
4.000
3.500
3.000
2.500
2.000
step applied
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
TIME IN MINUTES

back correction eliminated the off-set.

The experiments have shown that both methods of feedforward compensation are effective and are valid alternatives. The combined feedback-feedforward controller has made the process under study effectively immune to disturbances in feed conditions, and there appears to be no doubt that the method of feedforward compensation developed in this study can find applications in a variety of similar processes.

The feedforward compensation achieved by Wilson with analog components did not appear to produce results as satisfactory as the digital feedforward controller used here. All feedforward runs reported by Wilson contain fluctuations in the product response.

7.1.4 Inferential Control

The inferential controller has been used in a series of experiments labelled INF on Figures 7.2 and 7.3. This type of control is employed in situations where it is desirable to control a process variable without physically measuring it. It was the intention of this work to apply the concept to a process that was more complex than those described in the literature (24, 14). The results obtained yielded more information about the usefulness and flexibility of inferential control than had been anticipated.

It had been observed from open loop experiments

shown on Figures 7.5 and 7.6 that the model responded seven to eight minutes faster than the process. On implementing the inferential controller, provisions had been been made for arbitrary selection of the length of time that would elapse before the information obtained from the model was supplied to the controller. The four experiments of the series INF clearly show the effect of varying the time correspondence between model and process.

In runs INFI2 and INFII2, the controller obtained its information from a model that contained a time delay term which improved the agreement between the open loop response curves of the process and the model. In runs INFI and INFII, the information came from a model that did not contain the time delay and thus responded seven minutes faster than the process. The influence of this time difference is apparent from the diagrams. If allowance is made for the slight offset that actually did occur in the case of INFII, it is apparent that the undelayed model returned the process to its original steady state, whereas the delayed model did not. This result was not expected, nor can it be explained theoretically, since a time delay usually affects the speed of response of a given system, but not the final accuracy of the response curve when compared to the setpoint. Since the delay applied to the model was the only known difference between experiments INFI and INFII on the one hand and INFI2 and INFII2 on the other, no satisfactory explanation is

available for the off-set that resulted. As far as the time delay itself is concerned, it is natural to raise the question of the optimal time difference between model and process responses. It is felt that the limited number of experiments using the inferential controller have helped illustrate the importance of this question, rather than answer it. In this particular case, the controller obtained its information from the simplified model which did not contain a time delay term, but experimental runs had shown that it anticipated the process by approximately seven minutes. It is hoped that the results obtained will generate further work on this aspect of inferential control.

Returning to experiment INFII, it is felt that the slight off-set that occurred at the end of the run is a consequence of the model limitations. A cross reference to Figure A-16 of Appendix A shows that the model had returned to steady state. As is seen from the open loop run on Figure 7.6, the model did not exactly reproduce the process response. On programming the controller, the process model has been included as a replaceable unit, such that in future work more complete models can be substituted.

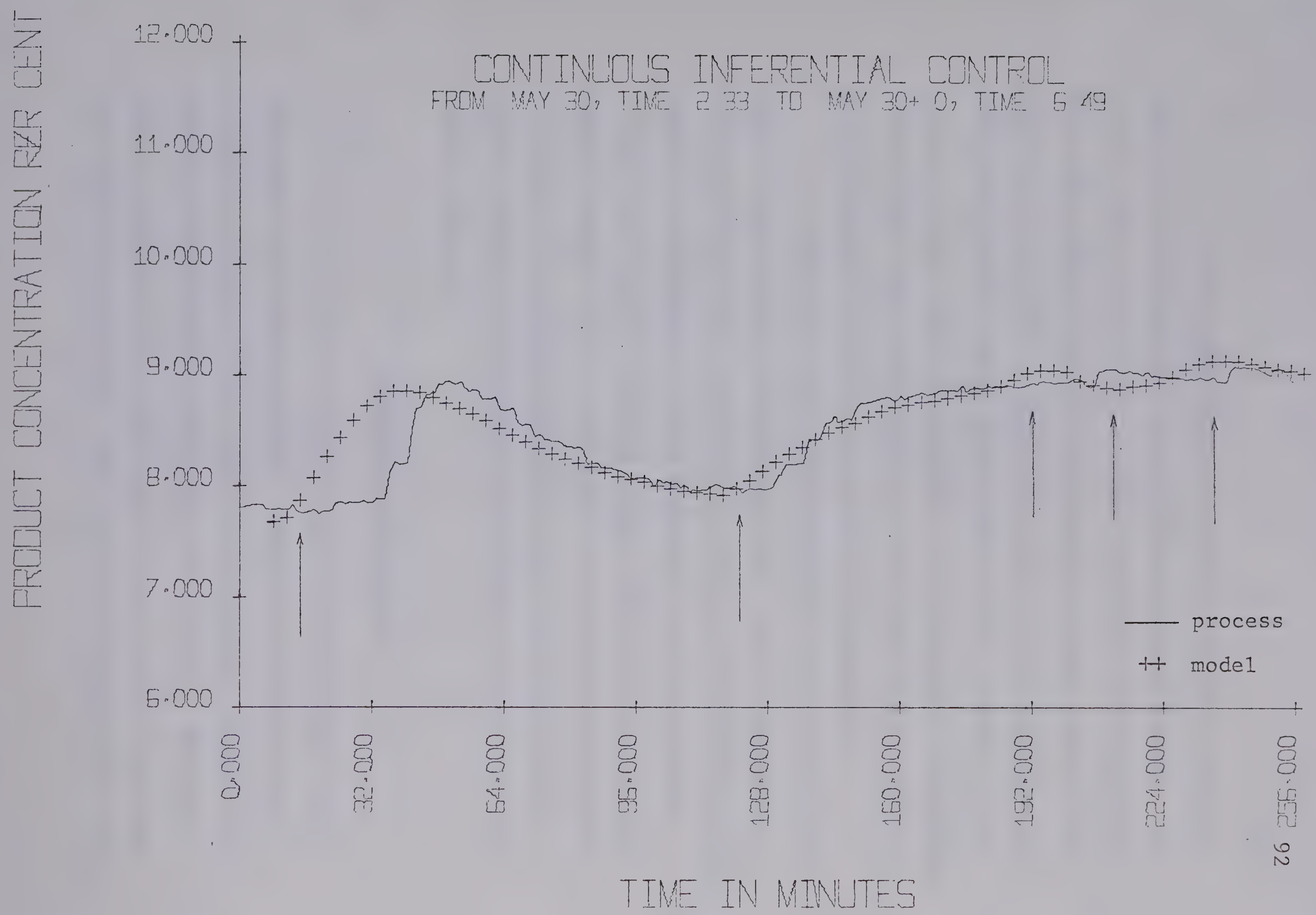
The question obviously will arise as to the relation between the accuracy of a model when compared to an open loop process response, and the quality of control that can be achieved with it in an inferential control system. One phenomenon that has become apparent in this work is that

although the model response anticipated the process and thus did not reproduce the open loop process response curve, the inferential control runs INFI and INFII resulted in better product quality control than the conventional feedback runs FBI3 and FBII2. The lead of the model with respect to the process is suggested as explanation for this phenomenon.

It is apparent that in an inferential control system, model parameters become in a sense control parameters, and it does not follow that a model that exactly reproduces the open loop process response is also the best one for control purposes.

On one occasion during the experimental program, the process was operated under inferential control for a period in excess of four hours. During this period, various disturbances were introduced into the process. The results are shown on Figure 7.9. The arrows indicate the instances at which step changes were applied. For the first run, inferential feedback alone was used. At the end of this run, feed-forward compensation was added. The step changes were the following:

1. feed flow 5.5 to 5.0 lbs/min and feed concentration 3.0 to 3.3%
2. setpoint 7.8 to 9.0%
3. feed concentration 3.3 to 3.6%
4. feed flow 5.0 to 5.3 lbs/min
5. feed flow 5.3 to 5.0 lbs/min



CONTINUOUS INFERENTIAL CONTROL
FROM MAY 30, TIME 2 33 TO MAY 30+ 0, TIME 6 49
PRODUCT CONCENTRATION PER CENT
12.000
11.000
10.000
9.000
8.000
7.000
6.000
0.000
32.000
64.000
96.000
128.000
160.000
192.000
224.000
256.000
TIME IN MINUTES
process
model

At the beginning of the first run, the refractometer did not function, as can be seen on the diagram. As soon as this was noticed, it was corrected, which explains the sudden steep rise of the process response curve. The malfunctioning did not have any effect on the operation in this case, as the model output was used for control. The process output was only recorded. The occasional deviations of the model from the process in the later part of the extended run are all within the error of measurement.

This experiment was intended to test the performance of the combined controller under a set of operating conditions that was realistic in the general industrial sense; namely, an extended run with frequent load disturbances, without necessarily reaching steady state between two successive changes. It is felt that the controller performed satisfactorily, which admits the belief that the control system developed is generally applicable to processes of a similar nature.

7.2 Equipment Performance

The process equipment performed satisfactorily throughout the entire program. Andre had reported that frequent switches of feed tanks were necessary which introduced undesired disturbances into the process. With the ratio controlled feed system used in this study, this type of problem did not arise. The product was pumped back into

the feed tank and feed concentration monitored by the ratio controller, such that the operation was continuous.

Wilson had faced the problem of sticky valves on various occasions due to the properties of the sugar solutions he used. The triethylene glycol used in this work did not present difficulties of this nature.

The higher throughput rates reduced the hold-up time, such that generally, the process regained steady state faster than was the case with Andre and Wilson.

The steam economy generally achieved in industrial double effect evaporators is 1.6 (16). In this work, it ran from 1.5 to 1.7 which is one indication of realistic operating conditions.

The effect of the averaging level control is best illustrated by Figures 7.7 and 7.8. Due to the dampened response of the bottoms flows, disturbances propagated only gradually and control of product quality was facilitated.

The errors of closure for all material balances are summarized on Table 7.1. Errors are expressed as percentage of inlet flows. All calculations were carried out using raw data obtained from the process at the time of the experiments concerned. As a rule, the steady state data at the start and at the end of all experiments were provided by a list of instantaneous readings that was printed by a program written especially for this purpose. Only in a very few cases were the noise components contained in these in-

Experiment	Total Process		Solute Balance		First Effect		Second Effect	
Number	start	end	start	end	start	end	start	end
FBI2	5.3	5.8	2.8	0.6	3.0	1.2	3.8	6.6
FBI3	3.0	7.0	0.6	5.0	2.6	4.0	0.6	4.7
FBII2	5.9	6.5	0.8	5.5	2.2	7.4	5.5	-1.5
FBII3	6.5	6.9	3.7	6.3	3.8	6.0	4.1	1.7
FFI2	4.0	4.4	4.6	1.8	3.6	-2.2	0.6	9.2
FFI3	3.0	2.0	3.6	3.6	1.0	3.4	3.0	-2.3
FFII2	6.4	7.6	9.0	8.0	5.5	4.9	1.4	4.2
FFII3	7.4	5.3	4.2	10.8	4.0	3.1	5.5	3.7
FFIV2	6.2	4.9	7.4	4.2	-0.7	2.7	10.9	3.4
FFIV1	6.3	7.7	4.2	7.4	-0.3	2.6	10.5	8.2
FFBI	7.1	6.3	5.0	0.0	6.5	5.2	7.0	1.8
FFBI2	6.4	6.9	7.9	1.5	2.6	1.6	6.1	8.7
FFBII3	3.8	2.1	2.9	10.7	3.5	0.2	1.4	3.0
FFBII2	6.1	5.9	4.5	4.7	6.1	5.9	0.0	0.0
INFI2	3.4	3.0	5.1	0.7	4.0	3.0	1.0	0.0
INFII2	3.1	3.6	5.9	8.5	1.5	2.9	2.5	1.2
INFI	5.0	3.8	6.5	8.0	2.0	2.2	4.5	2.6
INFII	4.4	4.1	0.7	7.3	2.6	2.3	3.6	2.8
OLI2	1.8	6.8	1.7	4.0	2.2	3.2	0.6	5.7
OLII2	6.6	2.8	6.0	3.0	4.0	0.9	4.8	2.9
OLIV	7.5	6.4	7.1	3.6	4.8	3.9	3.9	3.9
CONINF	4.2	6.1	3.6	6.6	2.9	4.6	1.9	3.3

Table 7.1 Percentage Errors in Material Balance Calculations

Experiment Number	Steam Economy start	Steam Economy end	Heat Transfer Coef. First Effect start	Heat Transfer Coef. First Effect end	Heat Transfer Coef. Second Effect start	Heat Transfer Coef. Second Effect end
FBI2	1.70	1.68	667	665	493	384
FBI3	1.74	1.64	661	668	400	392
FBII2	1.57	1.52	634	721	352	423
FBII3	1.54	1.56	718	687	400	346
FFI2	1.66	1.72	666	707	393	382
FFI3	1.68	1.55	975	1005	444	514
FFII2	1.56	1.51	665	681	342	427
FFII3	1.55	1.65	902	870	497	453
FFIV2	1.64	1.57	780	885	485	464
FFIV1	1.58	1.64	902	886	551	462
FFBI	1.65	1.62	697	665	398	348
FFBI2	1.62	1.56	985	869	482	520
FFBII3	1.66	1.61	870	900	442	512
FFBII2	1.58	1.57	717	715	425	357
INFI2	1.68	1.62	992	991	396	475
INFII2	1.67	1.64	860	844	465	520
INFI	1.67	1.71	675	664	435	493
INFII	1.67	1.73	647	664	406	440
OLI2	1.70	1.61	660	663	470	470
OLII2	1.59	1.61	664	625	480	460
OLIV	1.63	1.62	634	628	371	370
CONINF	1.62	1.58	660	660	482	416
Averages	1.63		756		437	

Table 7.2 Steam Economies and Heat Transfer Coefficients

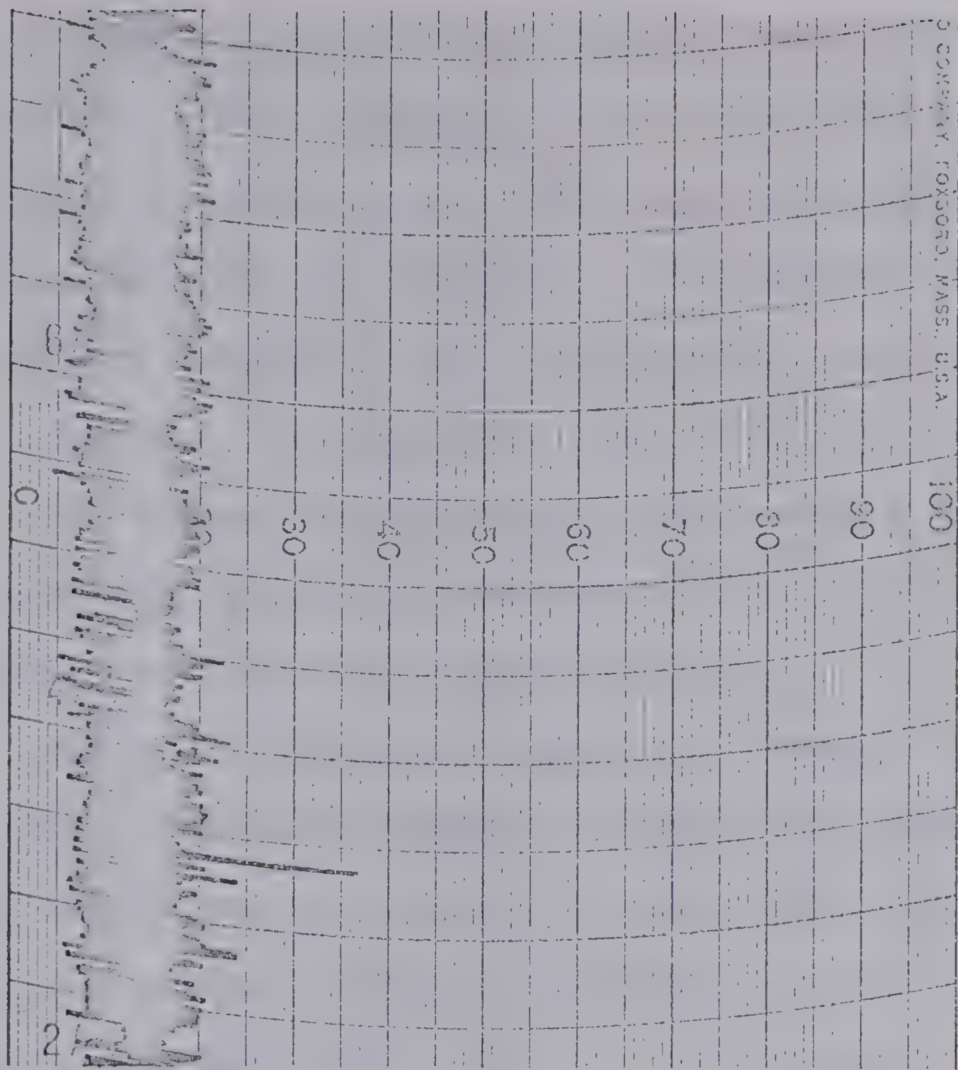


Figure 7.10 Recording of Condenser Condensate Flow F7

	first effect	second effect
Andre	379	358
Wilson	240	461
present study	756	437
handbook values(36)†	500	550

units are BTU/hr*ft^2*^{o}F

† The handbook values apply to similar industrially sized equipment and have only been included for illustrative purposes.

Table 7.3 Heat Transfer Coefficients

stantaneous readings so obvious that it was necessary to resort to the Foxboro recording charts for a more representative steady state value. This applied mainly to the condensate flow from the condenser. No control was exercised over this variable, and fluctuations were quite severe. A typical recording is shown on Figure 7.10.

The errors of closure for the overall balance were consistently positive, which must be attributed to slight shifts in instrument calibrations.

Table 7.2 shows the results of energy balance calculations. The heat transfer coefficients obtained for the first effect are considerably higher than those reported in previous studies. This is attributed to the higher throughput rates which gave rise to an increased heat flux and higher induced circulation velocities of the boiling liquid in the heating tubes of the first effect. The higher velocities are considered to be the main factor responsible for the increased heat transfer coefficients. In addition, the higher throughput rates resulted in pressures between 4 and 7 psig in the first effect. Andre and Wilson had operated at a slight vacuum. The boiling point of the solution in the first effect reported by Wilson was 190^{o}F. against 220^{o}F. for the present study. The higher boiling point resulted in a lower liquid viscosity which in turn gave rise to higher heat transfer coefficients (3). The average coefficient for the second effect was slightly lower than that reported by Wilson, which is attributed

to reduced circulation velocities that were necessary to obtain satisfactory level readings in the separator. A comparison of the heat transfer coefficients reported in various studies with this equipment is shown in Table 7.3.

The spread of the coefficient values for the first effect is larger than for the second effect. The experiments for which coefficients of 900 and over have been obtained were all conducted during a period when the vacuum service in the building did not supply the suction required for this process. It was necessary to lower the setpoint of the vacuum in order to maintain good control. This vacuum change carried through to the pressure in the first effect, amplified by a factor of two. The resulting boiling point changes were thus more pronounced in the first than in the second effect, such that the change in coefficient values was noticeable in the former case and not in the latter. All energy balances were closed by considering the heat losses to the surroundings, calculated by taking the difference between heat input and heat output. Average values of heat losses were as follows:

Total heat input		4960	BTU/min
Losses:	First effect	208	" "
	second effect	179	" "
	interstage piping	129	" "
	condenser	43	" "

Total losses 559 BTU/min or 11% of total heat input.

7.3 Direct Digital Control Performance

At the outset of this study it was mentioned that presently only 25% of digital computers installed in the chemical industry perform direct digital control. It is hoped that the experimental work presented here will contribute its modest share to demonstrate the versatility of direct digital control as applied to continuous processes.

In the present study, various complex configurations, such as cascade loops and ratio control loops, were applied successfully to the pilot plant scale evaporator. The advantages of these configurations have been demonstrated in the feed system and in the level controls used. There is no doubt that they can be applied effectively in other industrial installations as well.

The time sharing operating system of the computer allowed for the process to be controlled independently from other computer users. With the exception of some problems in the intial phase of the program, no difficulties were encountered as a result of the time-shared operation.

Since all measurements and setpoints could be made available in engineering units from the process operator's console, a steady state of reference could easily be defined and achieved, and load changes could be introduced with ease and high accuracy.

Although many aspects of direct digital control

were considered in this work, its capabilities have not been completely made use of. Digital controller tuning has not been employed in its true sense, and the possibilities offered by digital filters have only been exploited in a few specific cases.

Last but not least, it is expected that the present work has contributed to the development of standard software for direct digital control applications. The methods used to implement the feedforward as well as the inferential control system are not restricted to this particular equipment, but can be employed generally.

8 CONCLUSIONS

1. The advantages of a digital computer over conventional analog control instruments as illustrated by this work include the following:

   Control configurations were easily modified.

   Controller constants were implemented with high accuracy and changed readily. A practically unlimited range of constants was available.

   User written programs communicated with the direct digital control program. This feature allowed for the application of special algorithms and methods.

   A wide range of control schemes was implemented by interconnection of individual control loops.

2. A method of implementing feedforward control using "standard" DDC programs was demonstrated. This specific method was only one out of several possibilities offered by this system. The standard DDC software permits accurate implementation of various kinds of feedforward models as well as other arithmetic and numerical manipulations. In the case of the evaporator, feedforward compensation was based optionally on feed flow or on first effect bottoms flow signals. The two schemes were equally effective and were shown to be valid alternatives. The digital feedforward controller essentially prevented load

disturbances from affecting the product quality, but continued need for feedback correction was indicated to correct for off-sets and undetected or unknown disturbances.

3. The physical modifications of the equipment, the averaging level control scheme employed, as well as more suitable controller constants contributed to the achievement of improved control under the basic feedback scheme as compared to previous studies.
4. With the aid of the digital computer, inferential control was applied to the evaporation process using a very simple mathematical model. The performance of the inferential controller was influenced significantly by applying artificial leads or lags to the model with respect to the process.

NOMENCLATURE

ADC	Analog-to-digital converter
B1	First effect bottoms flow
B2	Second effect bottoms (product) flow
CI	Feed concentration
CO	Product concentration
COS	Current output station
CPU	Central processing unit
C1	Concentration in first effect
C2	Concentration at inlet to second effect
DAC	Digital-to-analog converter
DDC	Direct digital control
DDC Loop	Process variable record
delta	Denotes an increment
H	Computation interval
H1	Liquid hold-up in first effect
H2	Liquid hold-up in second effect
I/O	Input-output
K	Constant
KI	Integral control constant
KP	Proportional control constant
ma	Milli-amperes
meas	Measurement
MPX	Multiplexer
neg	Negative

O1	First effect overhead flow
O2	Second effect overhead flow
PD	Proportional-derivative
PI	Proportional-integral
POC	Process operator's console
poll time	Time interval between two successive calls to a variable record
phase time	Time elapsed between the beginning of a poll time interval and the call to the variable record, phase time $<$ poll time
PVT	Process variable table
ring buffer	Data accumulation record
s	Laplace operator
setpt	setpoint
ss	Steady state
STEC	Steam economy
t	Time
TEG	Triethylene glycol
TSX	Time-sharing executive system
V-core	Variable core
WI	Feed flow rate
WST	Steam flow rate
τ	Time constant
X'	A prime denotes a perturbation variable
$\overline{X}$	A horizontal bar denotes a steady state value

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Appendix A

Steady State of Reference

Reference is made to Figures A-1, B-19, 3.1, and 3.2 for the location of measuring devices

Variable Name	Variable Value		Variable Description
Flow Rates			
F1	2.00	lbs/min	steam flow to first effect
F2	3.30	lbs/min	first effect bottoms flow
F5	1.70	lbs/min	first effect overhead flow
F6	1.66	lbs/min	product flow
F7	1.64	lbs/min	second effect overhead flow
F8	5.00	lbs/min	total feed flow
F9	48.00	lbs/min	cooling water to condenser
F10	190.0	lbs/min	second effect circulation rate
F10	4.2	ft/sec	circulation velocity
Concentrations			
C1	3.2	percent	feed concentration
C6	9.65	percent	product concentration
Pressures			
P20	5.9	psig	first effect pressure
P22	-15.0	in Hg	second effect pressure
Levels			
L11	9.10	in H20	separator liquid level
L14	16.00	in H20	first effect liquid level
Temperatures			
T1	106.6	deg. F	condenser cooling water out
T2	227.0	deg. F	first effect vapour
T4	180.7	deg. F	solution to second effect
T5	246.1	deg. F	steam condensate first effect
T10	226.8	deg. F	steam to second effect
T11	123.0	deg. F	condenser condensate
T12	156.3	deg. F	separator vapour
T15	318.1	deg. F	steam supply to evaporator
T19	227.9	deg. F	liquid in first effect
T28	199.2	deg. F	steam condensate second effect
T29	65.0	deg. F	condenser cooling water in
T34	141.5	deg. F	product from second effect
T7	195.0	deg. F	feed to first effect

Table A-1

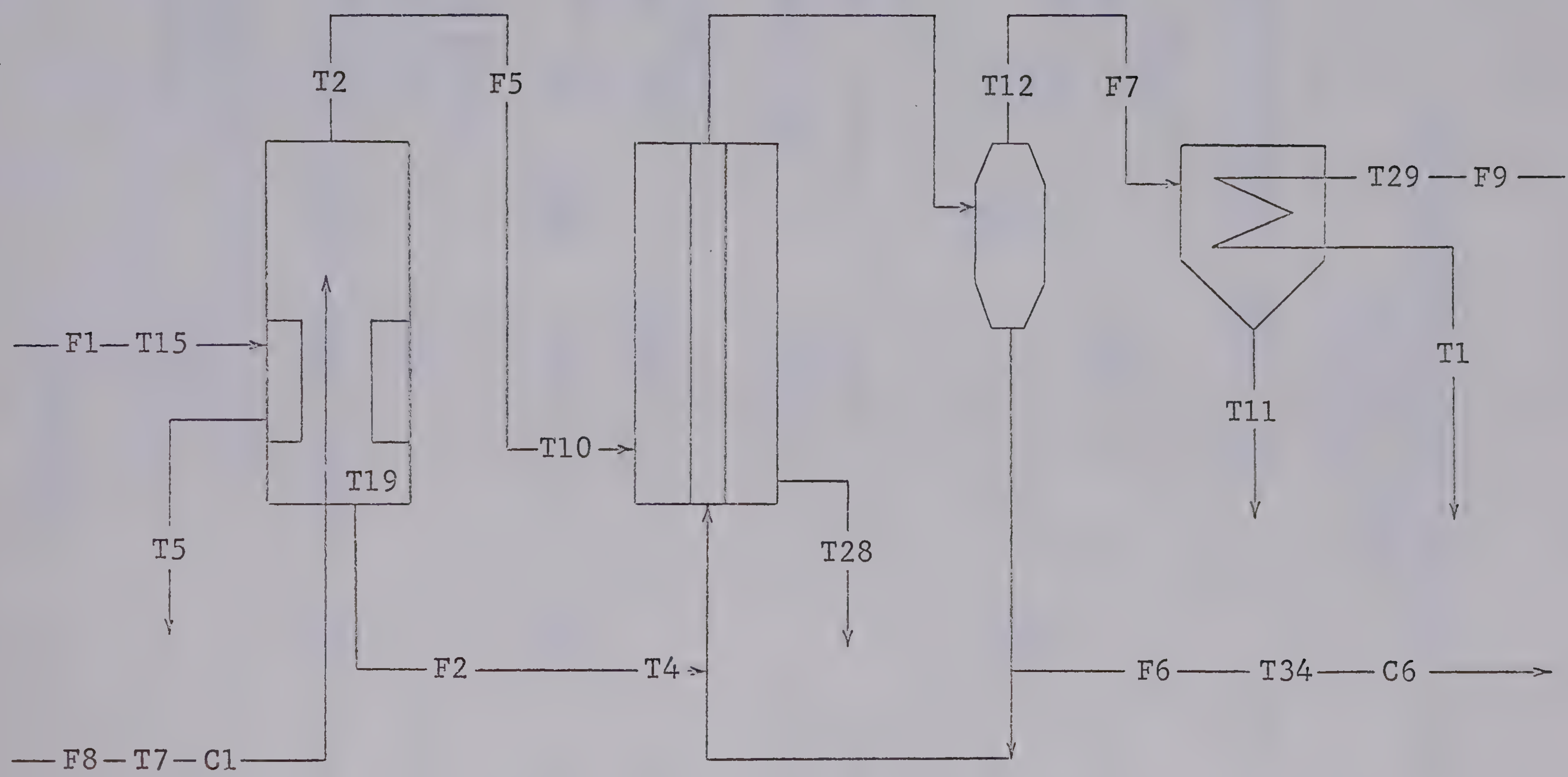


Figure A-1 Basis for Material and Energy Balance Calculations

Sample Material Balance, Experiment FBI2

Time Basis 1 Minute
Mass Units Lbs

Section Considered			Start	Error %		End	Error %
Total Process							
Input	F8		5.02			4.98	
Output	F5	1.73			1.46		
	F6	1.51			1.89		
	F7	1.51	4.75	5.3	1.34	4.69	5.8
Solute							
Input	F8*C1		0.140			0.169	
Output	F6*C6		0.136	2.8		0.170	0.6
First Effect							
Input	F8		5.02			4.98	
Output	F2	3.14			3.46		
	F5	1.73	4.87	3.0	1.46	4.92	1.2
Second Effect							
Input	F2		3.14			3.46	
Output	F6	1.51			1.89		
	F7	1.51	3.02	3.8	1.34	3.23	6.6
Steam Economy							
(F5+F6)/F1			1.70			1.68	

Table A-2

Sample Heat Balance, Experiment FBI2

Time Basis 1 Minute
Heat Units BTU

Section Considered		Stream	Start	Loss	End	Loss
First Effect	Input	F1 at T15	2250		1978	
		F8 at T7	792		794	
			3042		2772	
	Output	F1 at T5	384		334	
		F2 at T19	580		640	
		F5 at T2	2000		1680	
			2964	78	2654	118
Second Effect	Input	F5 at T10	2000		1680	
		F2 at T4	468		508	
			2468		2188	
	Output	F5 at T28	282		236	
		F6 at T34	222		284	
		F7 at T12	1705		1513	
			2209	259	2033	155
Condenser	Input	F7 at T12	1705		1513	
		F9 at T29	1773		1760	
			3478		3273	
	Output	F7 at T11	127		109	
		F9 at T1	3080		2960	
			3207	271	3069	204
Interstage Loss F2(T19-T4)				113		131

Heat Transfer Coefficients BTU/hr*^{o}F*ft^{2}

First Effect	$\dfrac{(F1atT15-F1atT5)*60}{9.88*(T5 - T19)}$	667	665
Second Effect	$\dfrac{(F5atT10-F5atT28)*60}{4.10*(T10 - T12)}$	493	384

Table A-3

Sample Heat Balance, Experiment FBI2

Time Basis 1 Minute
Heat Units BTU

Section Considered		Stream	Start	End
Total Process	Input	F1 at T15	2250	1978
		F8 at T7	792	794
		F9 at T29	1773	1760
			4815	4532
	Output	F1 at T5	384	334
		F5 at T28	282	236
		F6 at T34	222	284
		F7 at T11	127	109
		F9 at T1	3080	2960
			4095	3923
	Losses	First Effect	78	118
		Interstage	113	131
		Second Effect	259	155
		Condenser	271	204
			4816	4531
Balance Error %			0.0	0.0

Heat Transfer Areas

First Effect 9.88 ft^2

Second Effect 4.10 ft^2

Liquid Hold-up at Steady State

First Effect 23 lbs

Second Effect 28 lbs

Table A-3 continued

DATA FOR EXPERIMENT OL I 2

VARIABLE DESCRIPTION	VARI ABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.01	2.01
FIRST EFFECT BOTTOMS FLOW	F2	3.14	3.14
FIRST EFFECT OVERHEAD FLOW	F5	1.71	1.69
PRODUCT FLOW	F6	1.46	1.42
SECOND EFFECT OVERHEAD FLOW	F7	1.70	1.54
TOTAL FEED FLOW	F8	4.96	4.99
COOLING WATER TO CONDENSER	F9	43.03	43.12
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.40	3.00
PRODUCT CONCENTRATION	C6	8.00	10.08
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	103.00	103.30
FIRST EFFECT VAPOUR	T2	220.00	220.10
SOLUTION TO SECOND EFFECT	T4	181.40	182.80
STEAM CONDENSATE FIRST EFFECT	T5	238.30	238.70
FEED TO FIRST EFFECT	T7	189.40	190.40
STEAM TO SECOND EFFECT	T10	219.80	220.00
CONDENSER CONDENSATE	T11	119.90	118.20
SEPARATOR VAPOUR	T12	160.50	162.40
STEAM SUPPLY TO EVAPORATOR	T15	313.00	312.80
LIQUID IN FIRST EFFECT	T19	220.04	220.70
STEAM CONDENSATE SECOND EFFECT	T28	194.30	195.00
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	207.80	208.30

TABLE A- 4

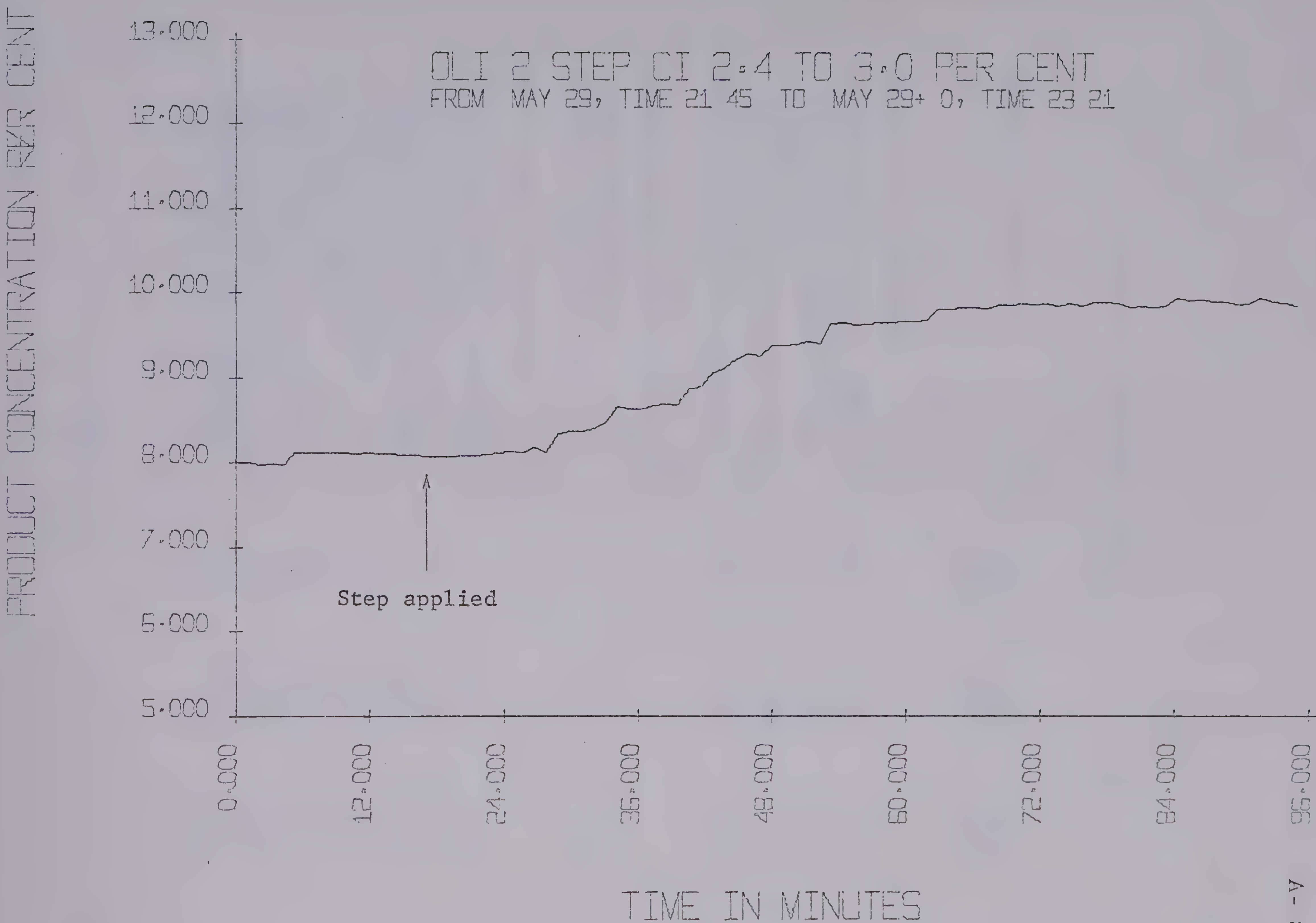

OLI 2 STEP CI 2.4 TO 3.0 PER CENT
FROM MAY 29, TIME 21 45 TO MAY 29+ 0, TIME 23 21
PRODUCT CONCENTRATION PER CENT
13.000
12.000
11.000
10.000
9.000
8.000
7.000
6.000
5.000
Step applied
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
TIME IN MINUTES

Figure A-2

DATA FOR EXPERIMENT FB I 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.90	1.67
FIRST EFFECT BOTTOMS FLOW	F2	3.14	3.46
FIRST EFFECT OVERHEAD FLOW	F5	1.73	1.46
PRODUCT FLOW	F6	1.51	1.89
SECOND EFFECT OVERHEAD FLOW	F7	1.51	1.34
TOTAL FEED FLOW	F8	5.02	4.98
COOLING WATER TO CONDENSER	F9	46.67	46.27
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.79	3.39
PRODUCT CONCENTRATION	C6	8.97	9.03
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	97.50	95.80
FIRST EFFECT VAPOUR	T2	216.50	215.70
SOLUTION TO SECOND EFFECT	T4	180.50	178.80
STEAM CONDENSATE FIRST EFFECT	T5	234.20	231.50
FEED TO FIRST EFFECT	T7	189.60	191.20
STEAM TO SECOND EFFECT	T10	216.40	215.80
CONDENSER CONDENSATE	T11	115.70	112.80
SEPARATOR VAPOUR	T12	159.20	155.30
STEAM SUPPLY TO EVAPORATOR	T15	313.70	314.80
LIQUID IN FIRST EFFECT	T19	217.30	216.60
STEAM CONDENSATE SECOND EFFECT	T28	195.30	193.50
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	178.70	181.90

TABLE A- 5

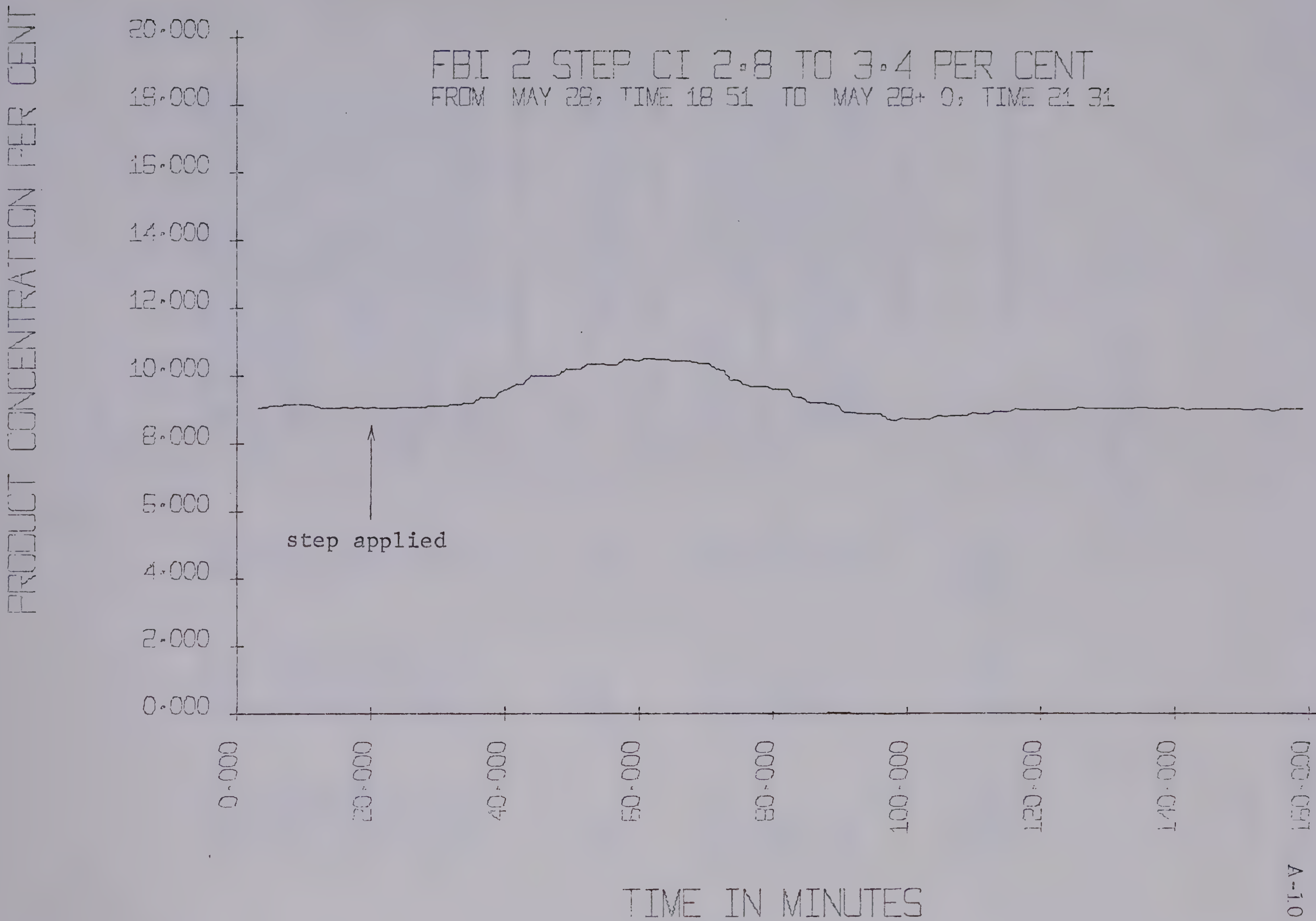
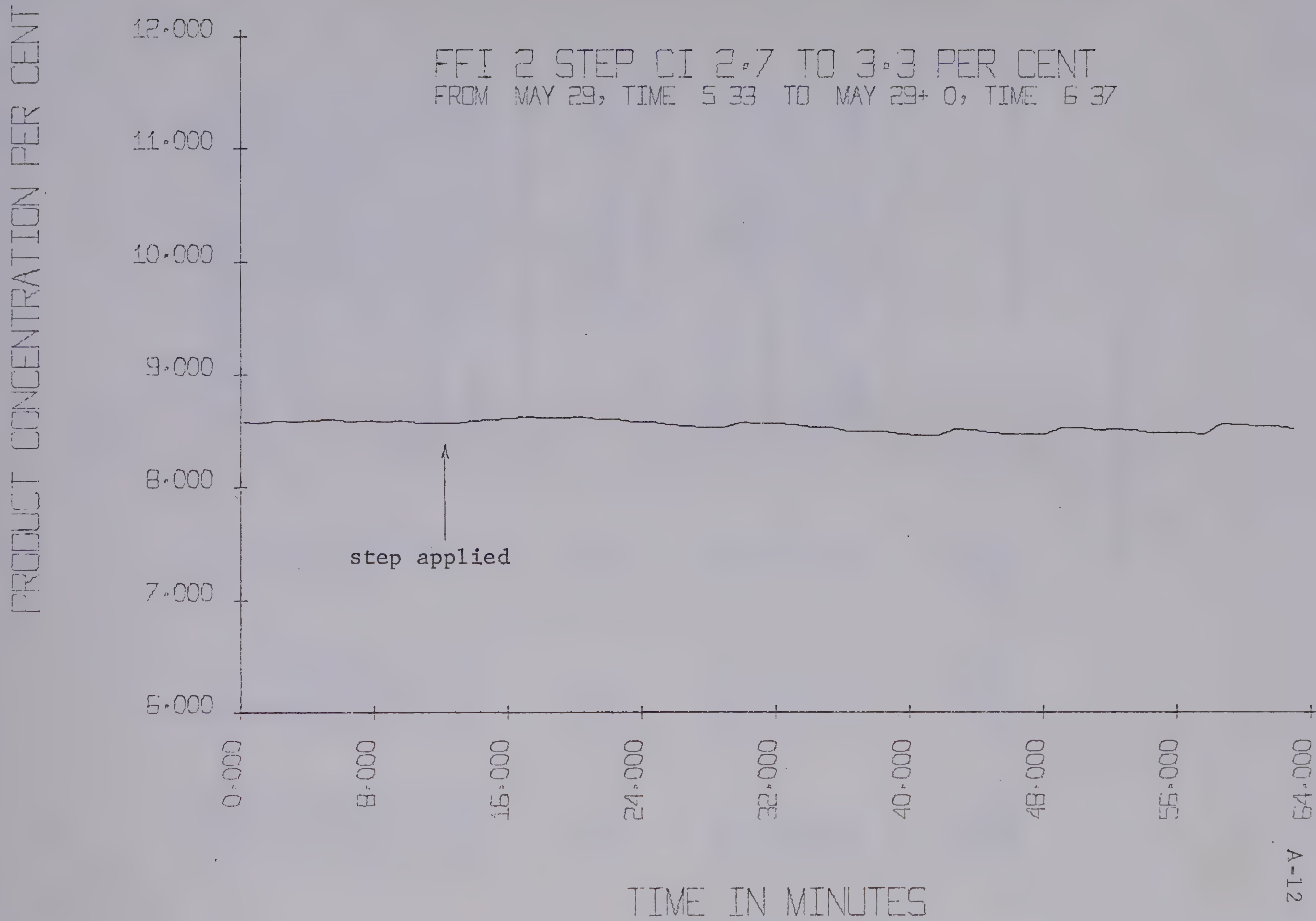

FBI 2 STEP CI 2·8 TO 3·4 PER CENT
FROM MAY 28, TIME 18 51 TO MAY 28+ 0, TIME 21 31
PRODUCT CONCENTRATION PER CENT
20·000
18·000
15·000
14·000
12·000
10·000
8·000
5·000
4·000
2·000
0·000
step applied
0·000
20·000
40·000
60·000
80·000
100·000
120·000
140·000
160·000
TIME IN MINUTES

Figure A- 3

DATA FOR EXPERIMENT FF I 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.90	1.65
FIRST EFFECT BOTTOMS FLOW	F2	3.16	3.56
FIRST EFFECT OVERHEAD FLOW	F5	1.64	1.52
PRODUCT FLOW	F6	1.63	1.92
SECOND EFFECT OVERHEAD FLOW	F7	1.51	1.31
TOTAL FEED FLOW	F8	4.98	4.97
COOLING WATER TO CONDENSER	F9	46.52	48.15
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.60	3.30
PRODUCT CONCENTRATION	C6	8.39	8.39
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	99.00	95.90
FIRST EFFECT VAPOUR	T2	215.20	210.60
SOLUTION TO SECOND EFFECT	T4	173.50	173.60
STEAM CONDENSATE FIRST EFFECT	T5	232.80	225.90
FEED TO FIRST EFFECT	T7	189.40	189.60
STEAM TO SECOND EFFECT	T10	215.10	210.70
CONDENSER CONDENSATE	T11	118.50	116.20
SEPARATOR VAPOUR	T12	146.80	146.00
STEAM SUPPLY TO EVAPORATOR	T15	312.30	313.90
LIQUID IN FIRST EFFECT	T19	215.70	211.50
STEAM CONDENSATE SECOND EFFECT	T28	187.20	189.00
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	192.70	193.80

TABLE A- 6



FFI 2 STEP CI 2.7 TO 3.3 PER CENT
FROM MAY 29, TIME 5 33 TO MAY 29+ 0, TIME 6 37
PRODUCT CONCENTRATION PER CENT
12.000
11.000
10.000
9.000
8.000
7.000
6.000
step applied
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
TIME IN MINUTES

Figure A- 4

DATA FOR EXPERIMENT FFB I

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.87	1.67
FIRST EFFECT BOTTOMS FLOW	F2	3.27	3.36
FIRST EFFECT OVERHEAD FLOW	F5	1.59	1.38
PRODUCT FLOW	F6	1.54	1.97
SECOND EFFECT OVERHEAD FLOW	F7	1.50	1.33
TOTAL FEED FLOW	F8	4.99	5.00
COOLING WATER TO CONDENSER	F9	47.09	46.94
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.79	3.39
PRODUCT CONCENTRATION	C6	8.49	8.60
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	99.40	96.10
FIRST EFFECT VAPOUR	T2	216.10	215.50
SOLUTION TO SECOND EFFECT	T4	175.70	175.80
STEAM CONDENSATE FIRST EFFECT	T5	233.40	231.10
FEED TO FIRST EFFECT	T7	190.10	190.90
STEAM TO SECOND EFFECT	T10	215.90	215.40
CONDENSER CONDENSATE	T11	119.90	116.00
SEPARATOR VAPOUR	T12	150.80	149.70
STEAM SUPPLY TO EVAPORATOR	T15	313.10	314.10
LIQUID IN FIRST EFFECT	T19	216.60	216.20
STEAM CONDENSATE SECOND EFFECT	T28	190.90	187.50
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	186.80	188.40

TABLE A- 7

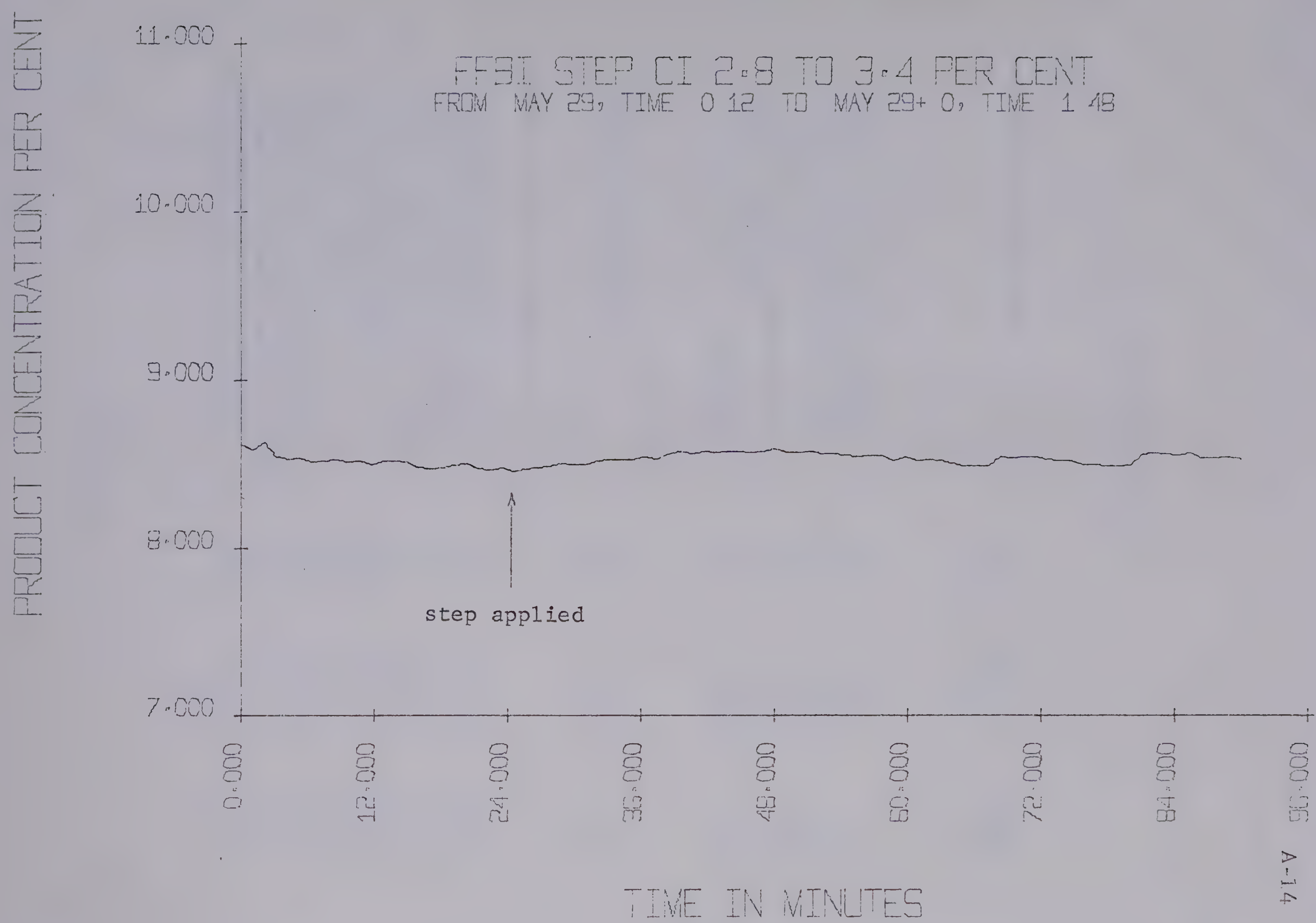

FFBI STEP CI 2·8 TO 3·4 PER CENT
FROM MAY 29, TIME 0 12 TO MAY 29+ 0, TIME 1 48
PRODUCT CONCENTRATION PER CENT
11·000
10·000
9·000
8·000
7·000
step applied
0·000
12·000
24·000
36·000
48·000
60·000
72·000
84·000
96·000
TIME IN MINUTES

Figure A- 5

DATA FOR EXPERIMENT FB I 3

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.66	1.90
FIRST EFFECT BOTTOMS FLOW	F2	3.29	3.18
FIRST EFFECT OVERHEAD FLOW	F5	1.58	1.61
PRODUCT FLOW	F6	1.96	1.53
SECOND EFFECT OVERHEAD FLOW	F7	1.31	1.50
TOTAL FEED FLOW	F8	5.00	4.99
COOLING WATER TO CONDENSER	F9	47.66	47.07
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.39	2.79
PRODUCT CONCENTRATION	C6	8.56	8.64
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	96.40	99.20
FIRST EFFECT VAPOUR	T2	215.30	214.20
SOLUTION TO SECOND EFFECT	T4	175.40	173.40
STEAM CONDENSATE FIRST EFFECT	T5	231.00	231.90
FEED TO FIRST EFFECT	T7	191.20	190.10
STEAM TO SECOND EFFECT	T10	215.30	214.20
CONDENSER CONDENSATE	T11	116.30	118.80
SEPARATOR VAPOUR	T12	149.60	147.40
STEAM SUPPLY TO EVAPORATOR	T15	314.00	312.50
LIQUID IN FIRST EFFECT	T19	216.20	214.90
STEAM CONDENSATE SECOND EFFECT	T28	182.70	189.50
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	188.80	191.90

TABLE A- 8

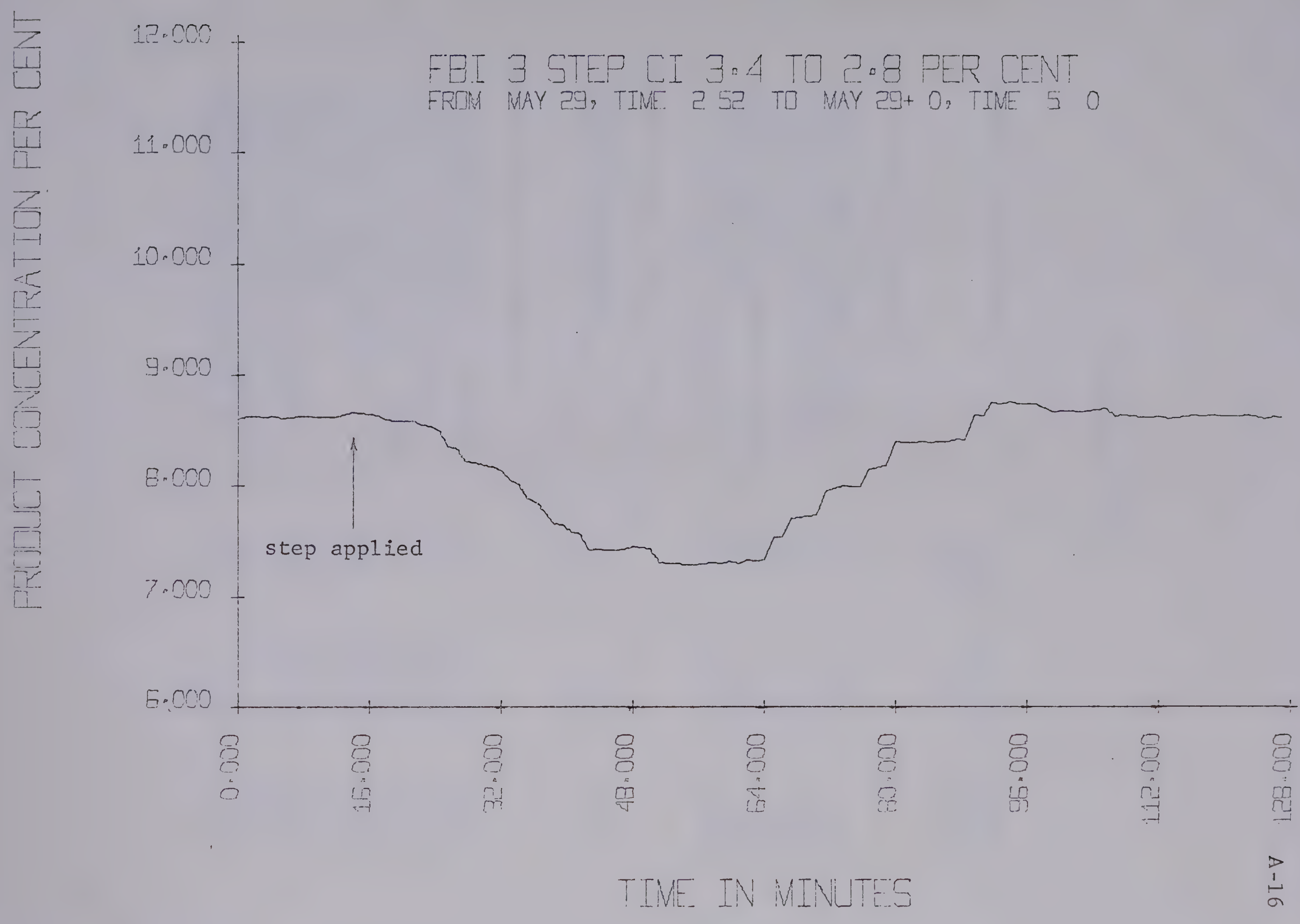

FBI 3 STEP CI 3.4 TO 2.8 PER CENT
FROM MAY 29, TIME 2 52 TO MAY 29+ 0, TIME 5 0
PRODUCT CONCENTRATION PER CENT
12.000
11.000
10.000
9.000
8.000
7.000
6.000
step applied
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
TIME IN MINUTES

Figure A- 6

DATA FOR EXPERIMENT FF I 3

VARIABLE DESCRIPTION	VARI ABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.82	2.22
FIRST EFFECT BOTTOMS FLOW	F2	3.35	2.96
FIRST EFFECT OVERHEAD FLOW	F5	1.55	1.87
PRODUCT FLOW	F6	1.75	1.46
SECOND EFFECT OVERHEAD FLOW	F7	1.50	1.57
TOTAL FEED FLOW	F8	4.95	5.00
COOLING WATER TO CONDENSER	F9	48.81	48.69
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.39	2.79
PRODUCT CONCENTRATION	C6	9.19	9.19
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	106.50	111.30
FIRST EFFECT VAPOUR	T2	229.00	231.10
SOLUTION TO SECOND EFFECT	T4	190.30	190.30
STEAM CONDENSATE FIRST EFFECT	T5	240.80	245.20
FEED TO FIRST EFFECT	T7	196.00	195.00
STEAM TO SECOND EFFECT	T10	228.70	230.90
CONDENSER CONDENSATE	T11	97.50	137.10
SEPARATOR VAPOUR	T12	172.50	173.10
STEAM SUPPLY TO EVAPORATOR	T15	316.40	314.20
LIQUID IN FIRST EFFECT	T19	229.50	231.70
STEAM CONDENSATE SECOND EFFECT	T28	207.60	209.80
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	126.20	123.60

TABLE A- 9

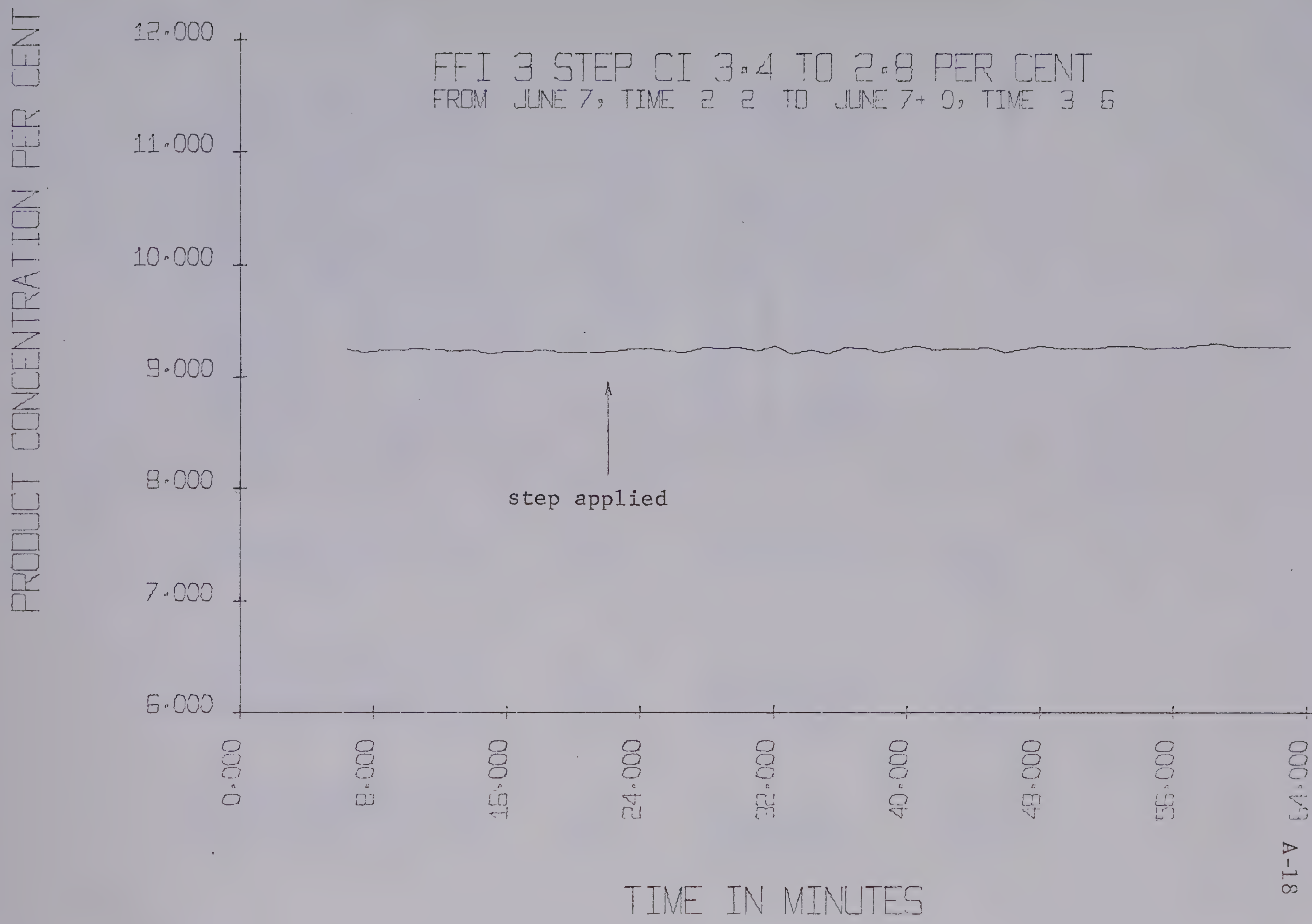

FFI 3 STEP CI 3.4 TO 2.8 PER CENT
FROM JUNE 7, TIME 2 2 TO JUNE 7+ 0, TIME 3 5
PRODUCT CONCENTRATION PER CENT
12.000
11.000
10.000
9.000
8.000
7.000
6.000
step applied
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
TIME IN MINUTES

Figure A- 7

DATA FOR EXPERIMENT FFB I 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.00	2.21
FIRST EFFECT BOTTOMS FLOW	F2	3.11	3.08
FIRST EFFECT OVERHEAD FLOW	F5	1.72	1.86
PRODUCT FLOW	F6	1.40	1.22
SECOND EFFECT OVERHEAD FLOW	F7	1.52	1.59
TOTAL FEED FLOW	F8	4.96	5.02
COOLING WATER TO CONDENSER	F9	49.33	57.64
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.30	2.70
PRODUCT CONCENTRATION	C6	10.77	10.86
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	109.30	110.30
FIRST EFFECT VAPOUR	T2	231.20	231.70
SOLUTION TO SECOND EFFECT	T4	191.00	191.30
STEAM CONDENSATE FIRST EFFECT	T5	244.40	246.60
FEED TO FIRST EFFECT	T7	196.20	195.20
STEAM TO SECOND EFFECT	T10	231.00	231.40
CONDENSER CONDENSATE	T11	105.60	133.50
SEPARATOR VAPOUR	T12	173.60	174.20
STEAM SUPPLY TO EVAPORATOR	T15	315.60	314.80
LIQUID IN FIRST EFFECT	T19	231.80	231.90
STEAM CONDENSATE SECOND EFFECT	T28	209.70	210.80
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	122.00	120.60

TABLE A-10

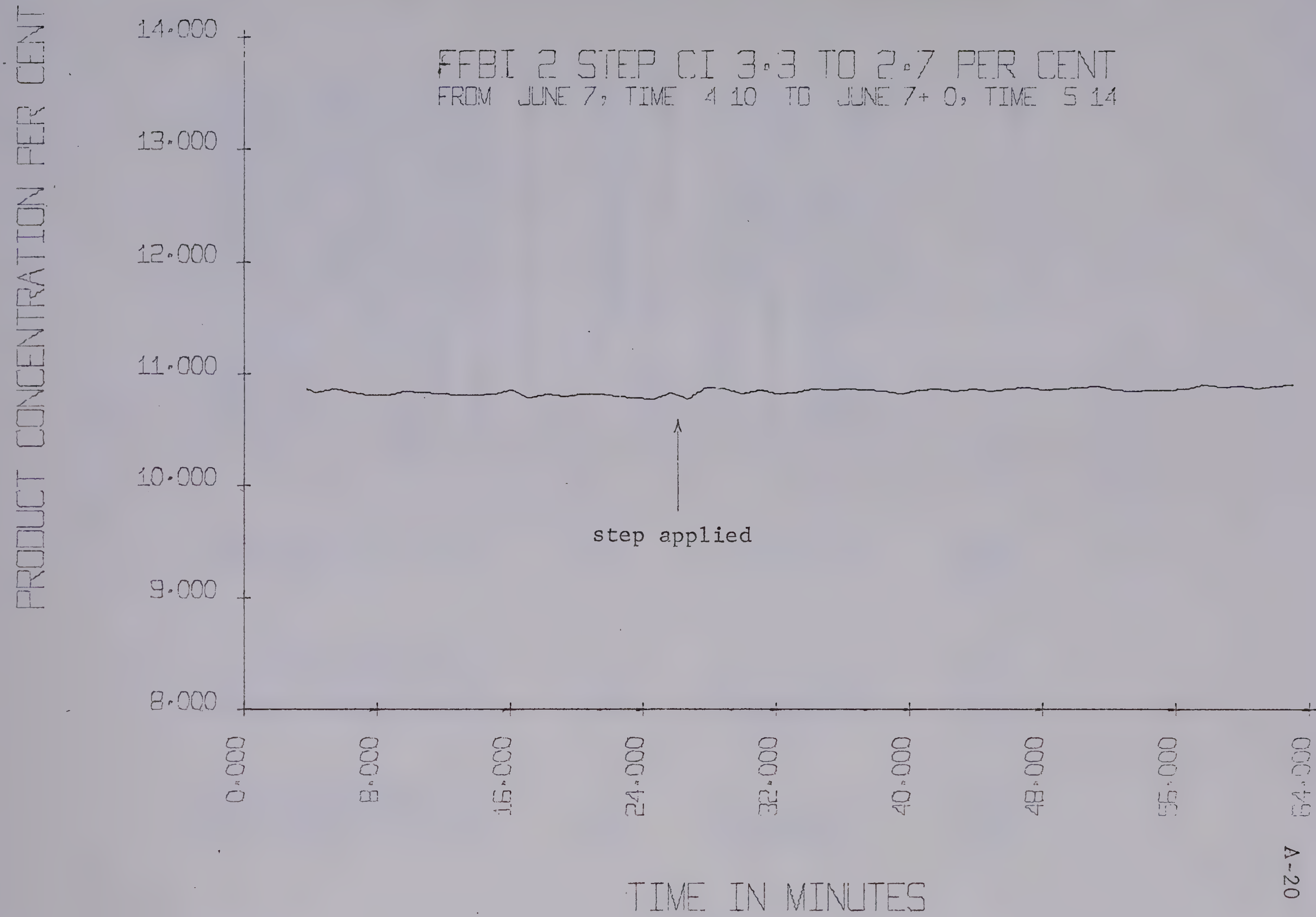

FFBI 2 STEP CI 3.3 TO 2.7 PER CENT
FROM JUNE 7, TIME 4 10 TO JUNE 7+ 0, TIME 5 14
PRODUCT CONCENTRATION PER CENT
14.000
13.000
12.000
11.000
10.000
9.000
8.000
step applied
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
TIME IN MINUTES

Figure A- 8

DATA FOR EXPERIMENT INF I

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.81	2.01
FIRST EFFECT BOTTOMS FLOW	F2	3.30	3.09
FIRST EFFECT OVERHEAD FLOW	F5	1.58	1.79
PRODUCT FLOW	F6	1.71	1.36
SECOND EFFECT OVERHEAD FLOW	F7	1.55	1.65
TOTAL FEED FLOW	F8	4.98	4.99
COOLING WATER TO CONDENSER	F9	42.91	43.17
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.09	2.50
PRODUCT CONCENTRATION	C6	8.30	8.31
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	99.20	102.50
FIRST EFFECT VAPOUR	T2	215.90	219.70
SOLUTION TO SECOND EFFECT	T4	179.30	180.80
STEAM CONDENSATE FIRST EFFECT	T5	232.40	237.80
FEED TO FIRST EFFECT	T7	189.90	189.70
STEAM TO SECOND EFFECT	T10	215.50	219.30
CONDENSER CONDENSATE	T11	115.50	120.10
SEPARATOR VAPOUR	T12	157.20	159.80
STEAM SUPPLY TO EVAPORATOR	T15	314.10	313.00
LIQUID IN FIRST EFFECT	T19	216.30	219.80
STEAM CONDENSATE SECOND EFFECT	T28	188.20	193.30
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	206.10	207.60

TABLE A-11

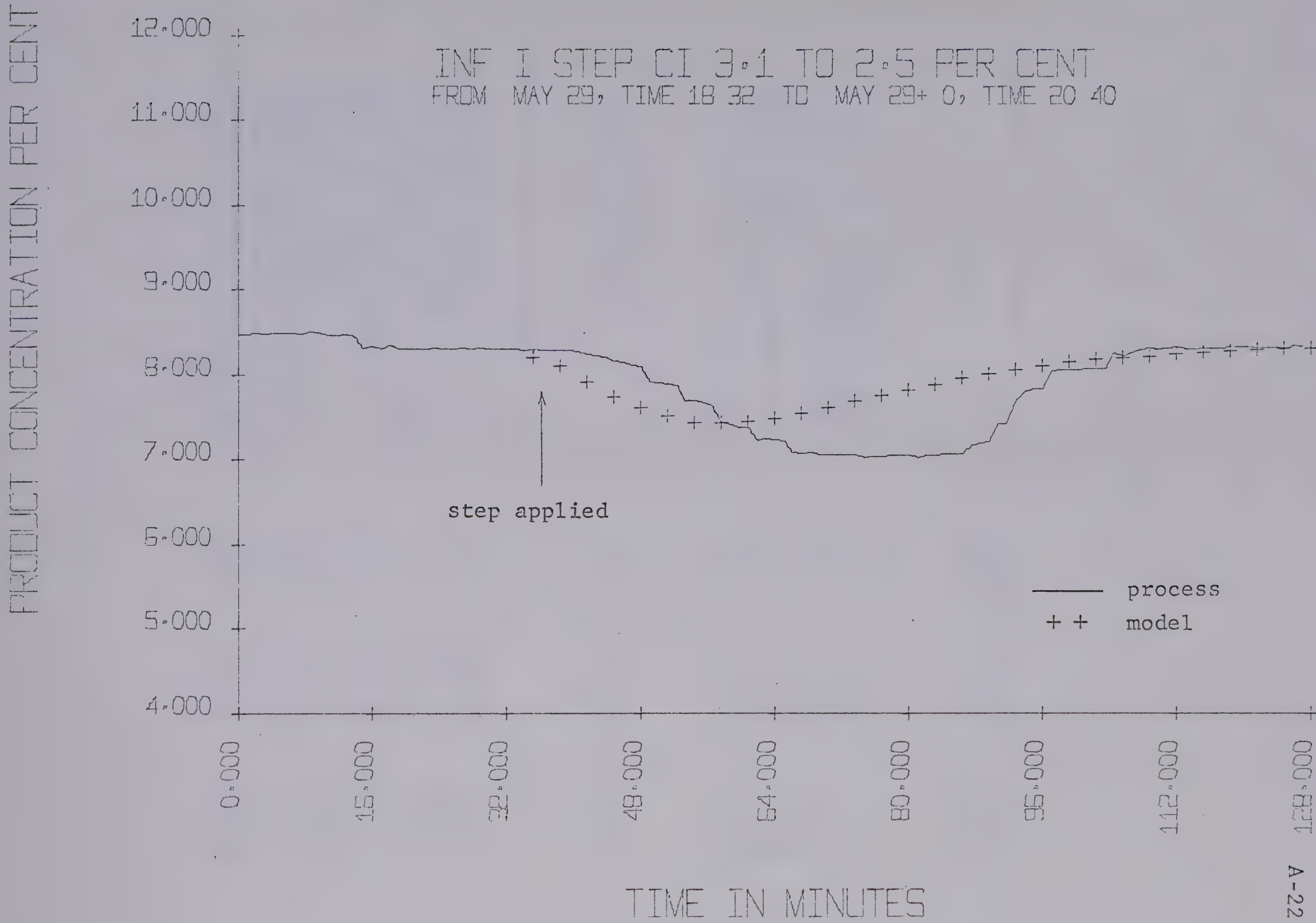

INF I STEP CI 3·1 TO 2·5 PER CENT
FROM MAY 29, TIME 18 32 TO MAY 29+ 0, TIME 20 40
PRODUCT CONCENTRATION PER CENT
12·000
11·000
10·000
9·000
8·000
7·000
6·000
5·000
4·000
0·000
16·000
32·000
48·000
64·000
80·000
96·000
112·000
128·000
TIME IN MINUTES
step applied
process
+ + model

Figure A- 9

DATA FOR EXPERIMENT INF I 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.85	2.02
FIRST EFFECT BOTTOMS FLOW	F2	3.21	3.11
FIRST EFFECT OVERHEAD FLOW	F5	1.57	1.76
PRODUCT FLOW	F6	1.71	1.60
SECOND EFFECT OVERHEAD FLOW	F7	1.53	1.51
TOTAL FEED FLOW	F8	4.98	5.02
COOLING WATER TO CONDENSER	F9	49.99	48.80
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.50	2.89
PRODUCT CONCENTRATION	C6	9.67	8.97
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	106.00	109.50
FIRST EFFECT VAPOUR	T2	229.40	231.10
SOLUTION TO SECOND EFFECT	T4	189.40	189.80
STEAM CONDENSATE FIRST EFFECT	T5	241.00	244.10
FEED TO FIRST EFFECT	T7	196.10	195.70
STEAM TO SECOND EFFECT	T10	229.00	230.90
CONDENSER CONDENSATE	T11	98.30	113.00
SEPARATOR VAPOUR	T12	171.70	172.30
STEAM SUPPLY TO EVAPORATOR	T15	316.20	315.50
LIQUID IN FIRST EFFECT	T19	229.70	231.60
STEAM CONDENSATE SECOND EFFECT	T28	207.00	209.10
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	127.40	127.50

TABLE A-12

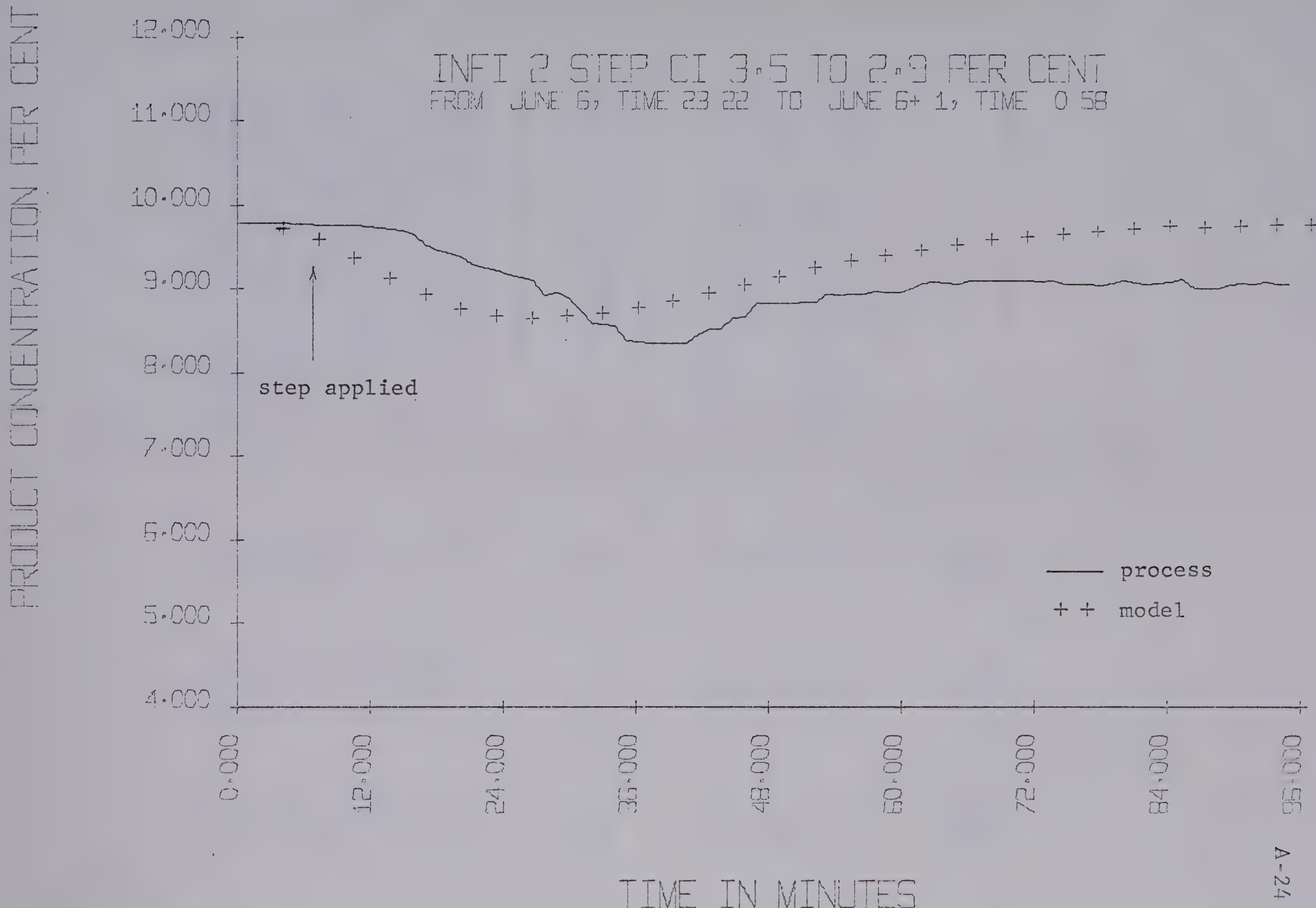

INFI 2 STEP CI 3.5 TO 2.9 PER CENT
FROM JUNE 6, TIME 23 22 TO JUNE 6+ 1, TIME 0 58
PRODUCT CONCENTRATION PER CENT
12.000
11.000
10.000
9.000
8.000
7.000
6.000
5.000
4.000
step applied
process
+ + model
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
TIME IN MINUTES

Figure A-10

DATA FOR EXPERIMENT OL II 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.01	2.01
FIRST EFFECT BOTTOMS FLOW	F2	3.13	3.78
FIRST EFFECT OVERHEAD FLOW	F5	1.70	1.65
PRODUCT FLOW	F6	1.42	2.09
SECOND EFFECT OVERHEAD FLOW	F7	1.51	1.55
TOTAL FEED FLOW	F8	5.00	5.50
COOLING WATER TO CONDENSER	F9	42.97	43.02
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.00
PRODUCT CONCENTRATION	C6	9.58	7.61
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	103.40	103.90
FIRST EFFECT VAPOUR	T2	220.40	222.90
SOLUTION TO SECOND EFFECT	T4	183.20	185.10
STEAM CONDENSATE FIRST EFFECT	T5	238.80	241.50
FEED TO FIRST EFFECT	T7	190.70	191.10
STEAM TO SECOND EFFECT	T10	220.20	222.70
CONDENSER CONDENSATE	T11	117.10	117.60
SEPARATOR VAPOUR	T12	163.20	164.20
STEAM SUPPLY TO EVAPORATOR	T15	313.00	312.60
LIQUID IN FIRST EFFECT	T19	220.90	223.40
STEAM CONDENSATE SECOND EFFECT	T28	195.80	197.10
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	208.60	208.90

TABLE A-13

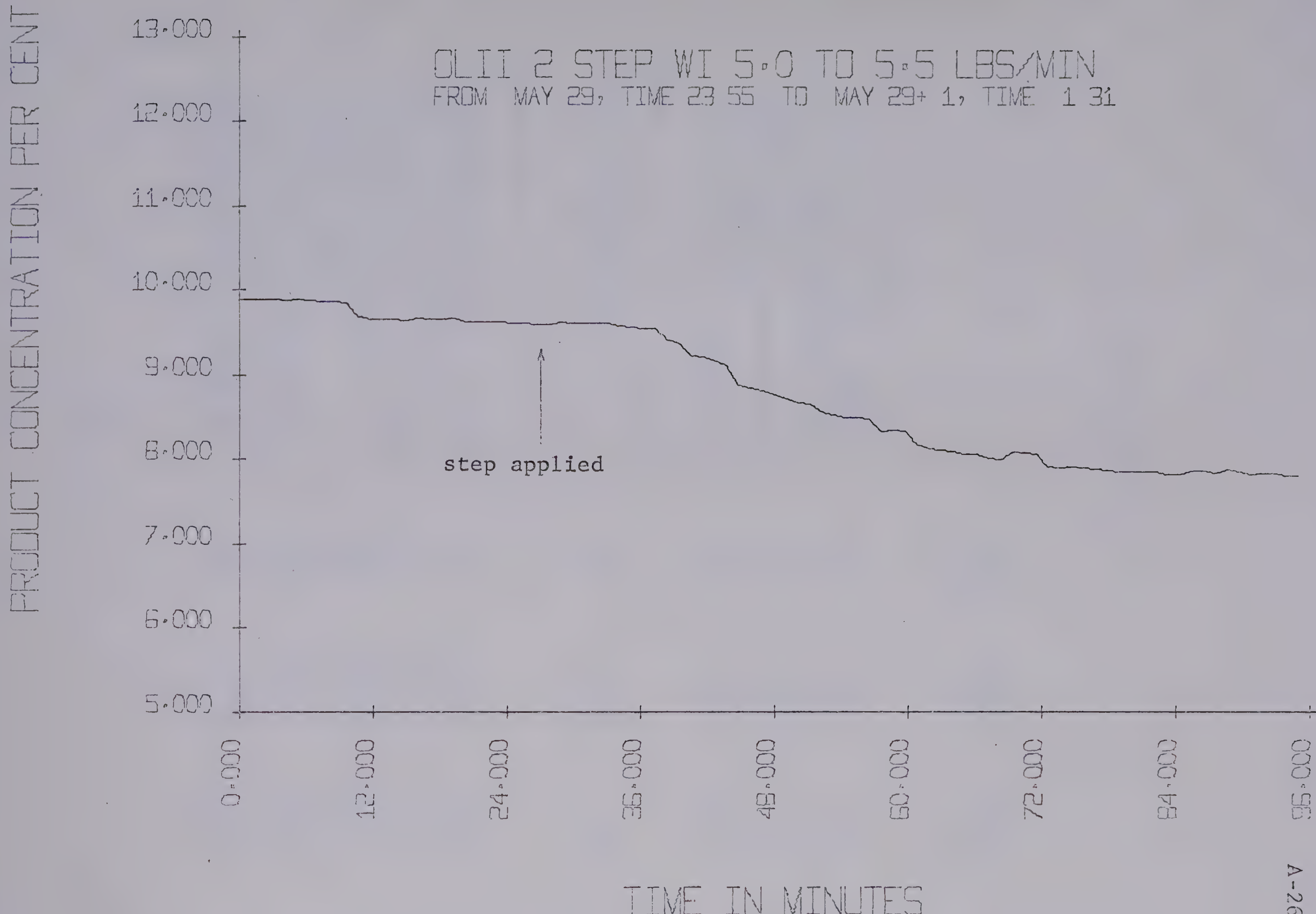

OLII 2 STEP WI 5.0 TO 5.5 LBS/MIN
FROM MAY 29, TIME 23 55 TO MAY 29+ 1, TIME 1 31
PRODUCT CONCENTRATION PER CENT
13.000
12.000
11.000
10.000
9.000
8.000
7.000
6.000
5.000
step applied
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
TIME IN MINUTES

Figure A-11

DATA FOR EXPERIMENT FB II 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.81	2.20
FIRST EFFECT BOTTOMS FLOW	F2	3.02	3.41
FIRST EFFECT OVERHEAD FLOW	F5	1.35	1.66
PRODUCT FLOW	F6	1.38	1.79
SECOND EFFECT OVERHEAD FLOW	F7	1.50	1.67
TOTAL FEED FLOW	F8	4.50	5.48
COOLING WATER TO CONDENSER	F9	48.35	48.78
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.80	2.80
PRODUCT CONCENTRATION	C6	9.04	9.11
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	96.20	101.60
FIRST EFFECT VAPOUR	T2	216.60	221.70
SOLUTION TO SECOND EFFECT	T4	178.40	181.40
STEAM CONDENSATE FIRST EFFECT	T5	233.60	241.40
FEED TO FIRST EFFECT	T7	189.20	189.90
STEAM TO SECOND EFFECT	T10	216.40	221.40
CONDENSER CONDENSATE	T11	114.50	121.80
SEPARATOR VAPOUR	T12	154.00	157.40
STEAM SUPPLY TO EVAPORATOR	T15	318.50	314.90
LIQUID IN FIRST EFFECT	T19	217.30	222.50
STEAM CONDENSATE SECOND EFFECT	T28	193.20	197.80
CONDENSER COOLING WATER IN	T29	65.00	65.00
PRODUCT FROM SECOND EFFECT	T34	195.70	196.00

TABLE A-14

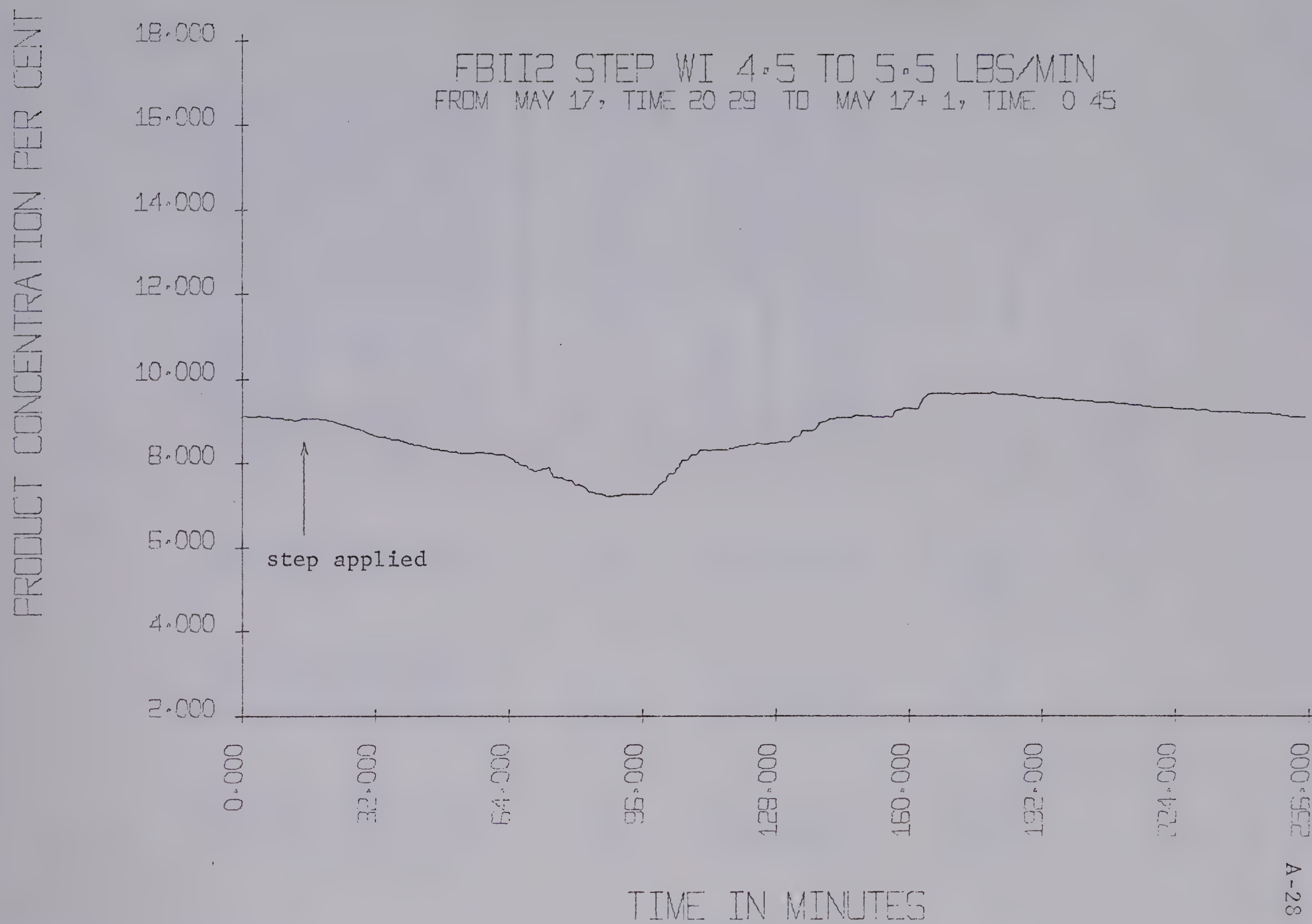

FBII2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM MAY 17, TIME 20 29 TO MAY 17+ 1, TIME 0 45
PRODUCT CONCENTRATION PER CENT
18.000
16.000
14.000
12.000
10.000
8.000
6.000
4.000
2.000
step applied
0.000
32.000
64.000
96.000
128.000
160.000
192.000
224.000
256.000
TIME IN MINUTES

Figure A-12

DATA FOR EXPERIMENT FF II 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.79	2.19
FIRST EFFECT BOTTOMS FLOW	F2	2.94	3.57
FIRST EFFECT OVERHEAD FLOW	F5	1.30	1.65
PRODUCT FLOW	F6	1.40	1.77
SECOND EFFECT OVERHEAD FLOW	F7	1.50	1.65
TOTAL FEED FLOW	F8	4.49	5.49
COOLING WATER TO CONDENSER	F9	49.22	48.91
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.69	2.69
PRODUCT CONCENTRATION	C6	9.49	9.13
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	98.40	103.70
FIRST EFFECT VAPOUR	T2	220.90	221.30
SOLUTION TO SECOND EFFECT	T4	182.00	183.10
STEAM CONDENSATE FIRST EFFECT	T5	237.50	241.00
FEED TO FIRST EFFECT	T7	189.20	190.00
STEAM TO SECOND EFFECT	T10	220.50	220.90
CONDENSER CONDENSATE	T11	109.60	123.30
SEPARATOR VAPOUR	T12	159.80	159.40
STEAM SUPPLY TO EVAPORATOR	T15	319.10	315.70
LIQUID IN FIRST EFFECT	T19	221.60	222.40
STEAM CONDENSATE SECOND EFFECT	T28	197.00	198.20
CONDENSER COOLING WATER IN	T29	65.00	65.00
PRODUCT FROM SECOND EFFECT	T34	198.20	198.10

TABLE A-15

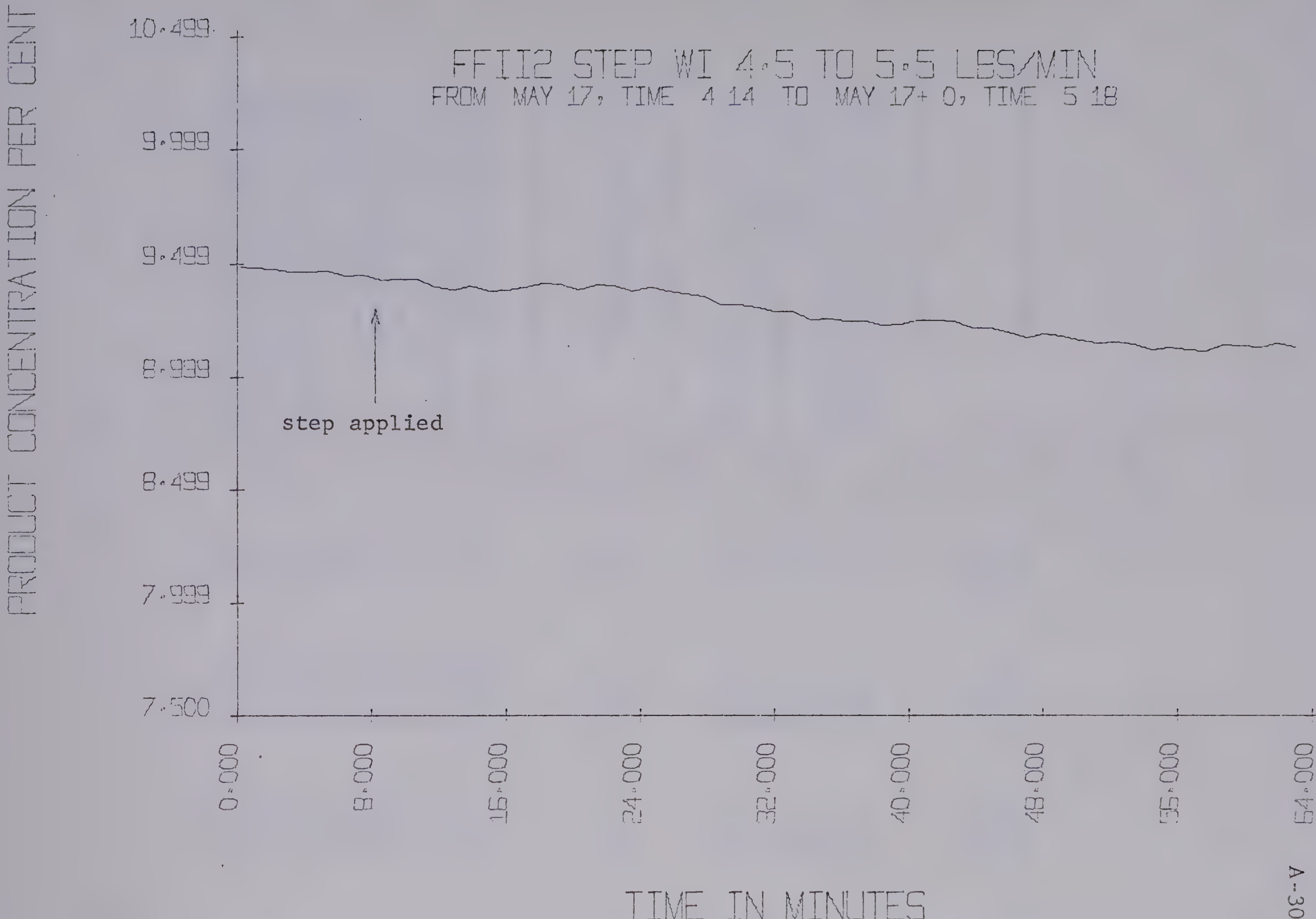

FFII2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM MAY 17, TIME 4 14 TO MAY 17+ 0, TIME 5 18
PRODUCT CONCENTRATION PER CENT
10.499
9.999
9.499
8.999
8.499
7.999
7.500
step applied
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
TIME IN MINUTES

Figure A-13

DATA FOR EXPERIMENT FF IV 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.84	2.26
FIRST EFFECT BOTTOMS FLOW	F2	2.81	3.52
FIRST EFFECT OVERHEAD FLOW	F5	1.72	1.85
PRODUCT FLOW	F6	1.20	1.70
SECOND EFFECT OVERHEAD FLOW	F7	1.30	1.70
TOTAL FEED FLOW	F8	4.50	5.52
COOLING WATER TO CONDENSER	F9	49.63	51.68
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.00
PRODUCT CONCENTRATION	C6	10.38	10.10
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	103.00	107.30
FIRST EFFECT VAPOUR	T2	228.90	235.00
SOLUTION TO SECOND EFFECT	T4	189.60	190.20
STEAM CONDENSATE FIRST EFFECT	T5	241.30	250.70
FEED TO FIRST EFFECT	T7	195.10	195.70
STEAM TO SECOND EFFECT	T10	228.60	234.80
CONDENSER CONDENSATE	T11	94.30	135.60
SEPARATOR VAPOUR	T12	171.80	171.10
STEAM SUPPLY TO EVAPORATOR	T15	317.60	313.20
LIQUID IN FIRST EFFECT	T19	229.40	236.00
STEAM CONDENSATE SECOND EFFECT	T28	207.10	212.30
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	90.00	84.70

TABLE A-16

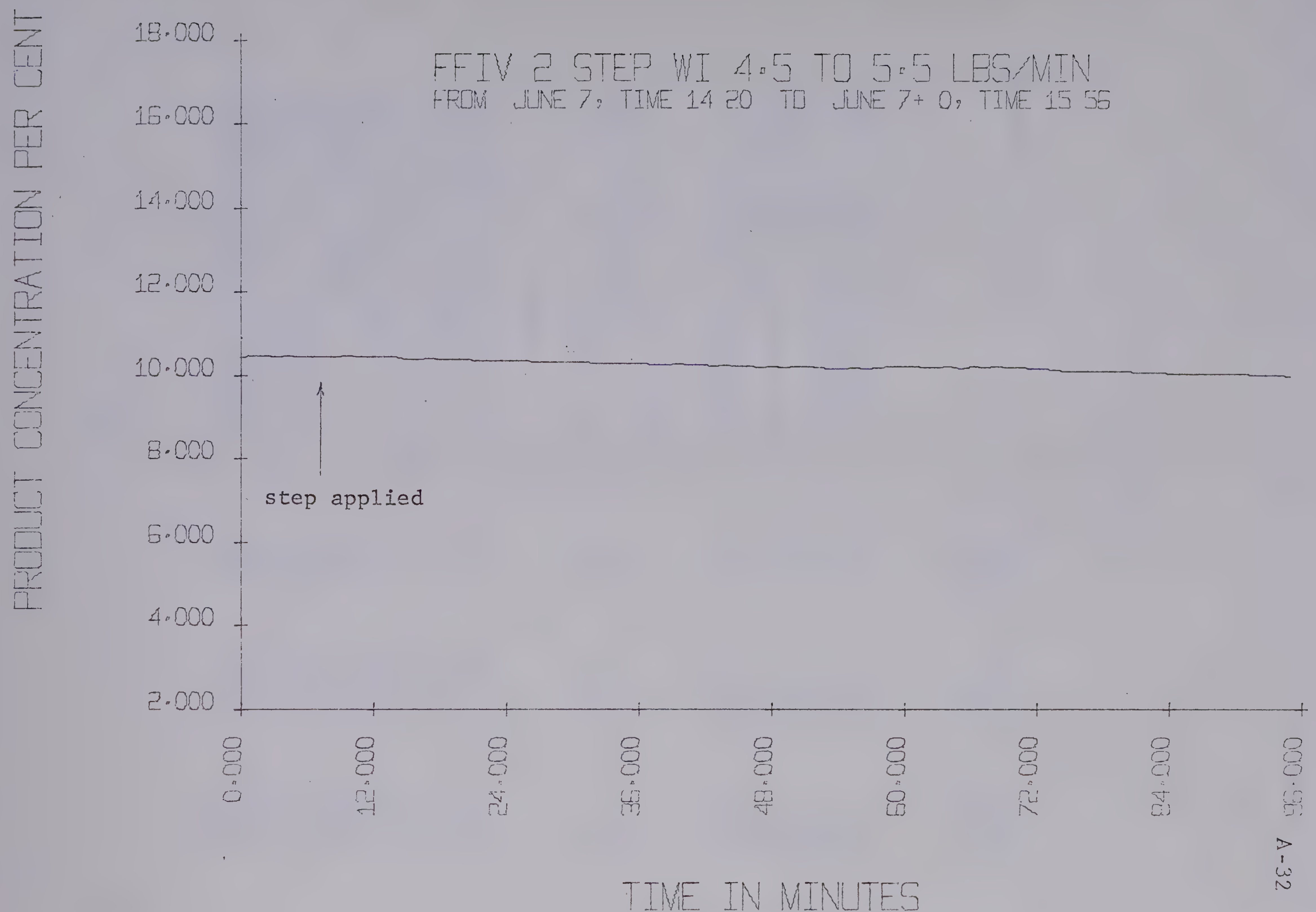

FFIV 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 14 20 TO JUNE 7+ 0, TIME 15 56
PRODUCT CONCENTRATION PER CENT
18.000
16.000
14.000
12.000
10.000
8.000
6.000
4.000
2.000
step applied
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
TIME IN MINUTES

Figure A-14

DATA FOR EXPERIMENT FFB II 3

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.92	2.30
FIRST EFFECT BOTTOMS FLOW	F2	2.72	3.61
FIRST EFFECT OVERHEAD FLOW	F5	1.63	1.89
PRODUCT FLOW	F6	1.16	1.69
SECOND EFFECT OVERHEAD FLOW	F7	1.55	1.81
TOTAL FEED FLOW	F8	4.51	5.51
COOLING WATER TO CONDENSER	F9	39.51	50.01
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.00
PRODUCT CONCENTRATION	C6	11.26	11.18
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	105.40	109.70
FIRST EFFECT VAPOUR	T2	232.20	231.90
SOLUTION TO SECOND EFFECT	T4	190.40	191.10
STEAM CONDENSATE FIRST EFFECT	T5	245.00	248.20
FEED TO FIRST EFFECT	T7	195.00	195.90
STEAM TO SECOND EFFECT	T10	231.70	231.60
CONDENSER CONDENSATE	T11	98.40	138.80
SEPARATOR VAPOUR	T12	173.20	173.00
STEAM SUPPLY TO EVAPORATOR	T15	316.60	313.20
LIQUID IN FIRST EFFECT	T19	232.50	232.90
STEAM CONDENSATE SECOND EFFECT	T28	210.20	210.90
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	97.10	95.20

TABLE A-17

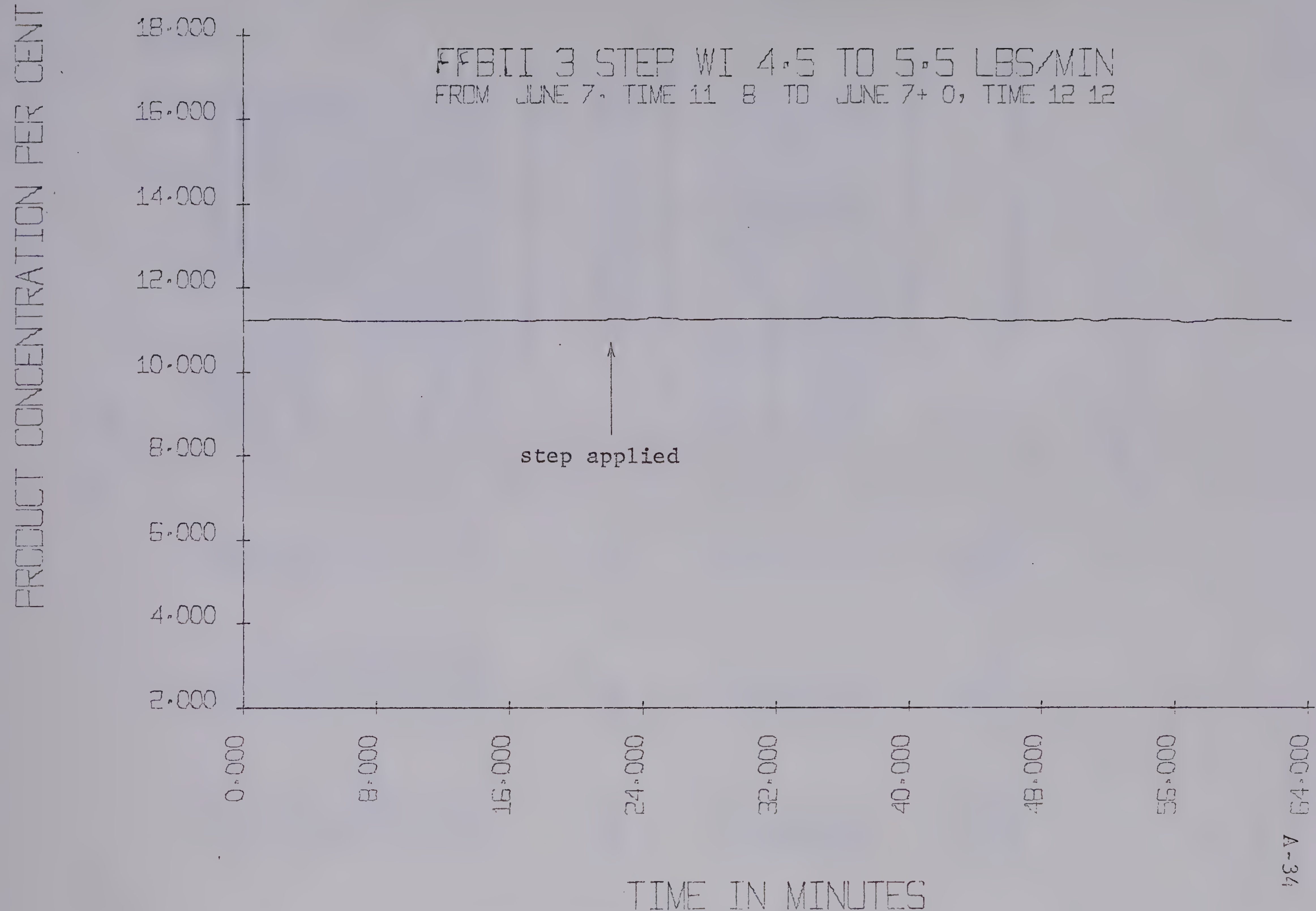

FFBII 3 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 11 8 TO JUNE 7+ 0, TIME 12 12
PRODUCT CONCENTRATION PER CENT
18.000
16.000
14.000
12.000
10.000
8.000
6.000
4.000
2.000
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
step applied
TIME IN MINUTES

Figure A-15

DATA FOR EXPERIMENT INF II

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.85	2.24
FIRST EFFECT BOTTOMS FLOW	F2	2.75	3.47
FIRST EFFECT OVERHEAD FLOW	F5	1.62	1.88
PRODUCT FLOW	F6	1.21	1.38
SECOND EFFECT OVERHEAD FLOW	F7	1.46	1.99
TOTAL FEED FLOW	F8	4.49	5.48
COOLING WATER TO CONDENSER	F9	47.36	47.39
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.00
PRODUCT CONCENTRATION	C6	10.98	11.01
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	101.60	106.20
FIRST EFFECT VAPOUR	T2	217.90	217.80
SOLUTION TO SECOND EFFECT	T4	177.40	176.30
STEAM CONDENSATE FIRST EFFECT	T5	235.90	239.10
FEED TO FIRST EFFECT	T7	189.40	190.90
STEAM TO SECOND EFFECT	T10	218.00	217.70
CONDENSER CONDENSATE	T11	121.60	121.80
SEPARATOR VAPOUR	T12	152.80	149.30
STEAM SUPPLY TO EVAPORATOR	T15	317.00	311.80
LIQUID IN FIRST EFFECT	T19	219.00	219.00
STEAM CONDENSATE SECOND EFFECT	T28	193.20	193.50
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	154.70	156.00

TABLE A-18

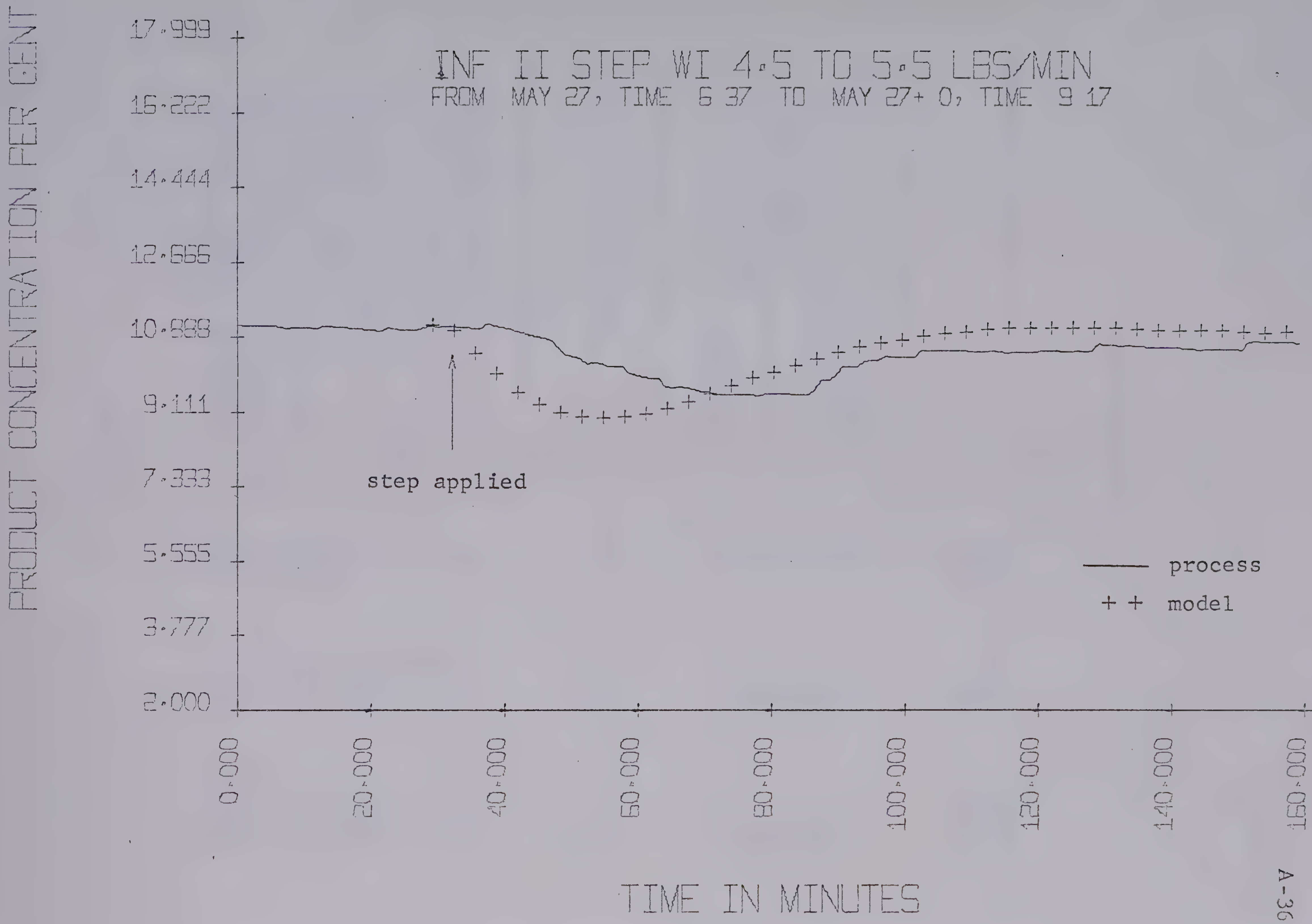

INF II STEP WI 4.5 TO 5.5 LBS/MIN
FROM MAY 27, TIME 6 37 TO MAY 27+ 0, TIME 9 17
PRODUCT CONCENTRATION PER CENT
17.999
16.222
14.444
12.666
10.888
9.111
7.333
5.555
3.777
2.000
0.000
20.000
40.000
60.000
80.000
100.000
120.000
140.000
160.000
TIME IN MINUTES
step applied
process
+ + model

Figure A-16

DATA FOR EXPERIMENT INF II 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.90	2.31
FIRST EFFECT BOTTOMS FLOW	F2	2.81	3.44
FIRST EFFECT OVERHEAD FLOW	F5	1.63	1.88
PRODUCT FLOW	F6	1.19	1.50
SECOND EFFECT OVERHEAD FLOW	F7	1.55	1.90
TOTAL FEED FLOW	F8	4.51	5.48
COOLING WATER TO CONDENSER	F9	50.03	45.66
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.00
PRODUCT CONCENTRATION	C6	10.77	9.96
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	106.70	115.70
FIRST EFFECT VAPOUR	T2	231.60	232.10
SOLUTION TO SECOND EFFECT	T4	191.60	191.60
STEAM CONDENSATE FIRST EFFECT	T5	245.10	248.60
FEED TO FIRST EFFECT	T7	195.20	195.80
STEAM TO SECOND EFFECT	T10	231.30	231.90
CONDENSER CONDENSATE	T11	97.30	143.20
SEPARATOR VAPOUR	T12	174.70	174.20
STEAM SUPPLY TO EVAPORATOR	T15	317.30	313.30
LIQUID IN FIRST EFFECT	T19	231.80	232.60
STEAM CONDENSATE SECOND EFFECT	T28	210.10	211.10
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	115.00	107.90

TABLE A-19

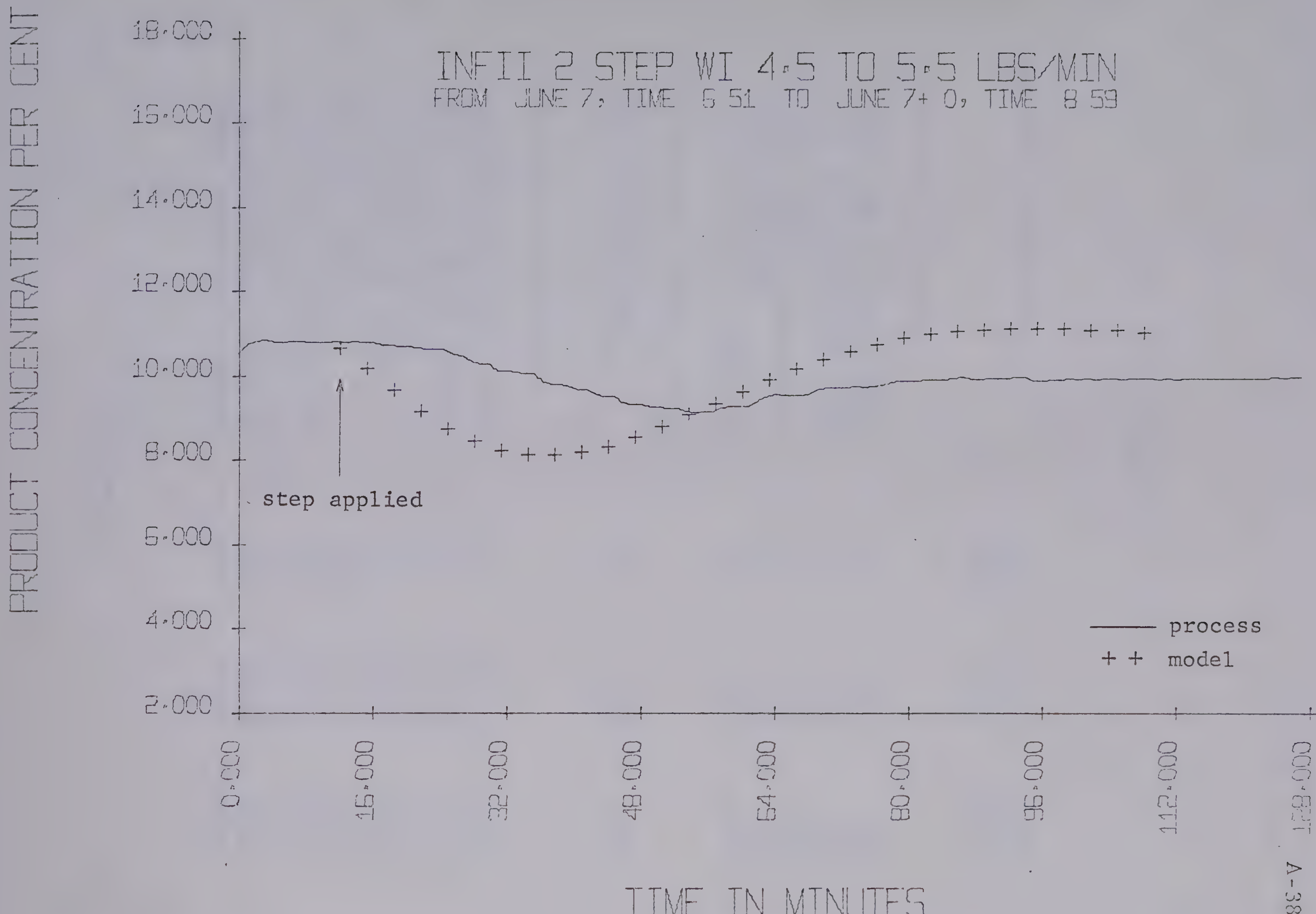
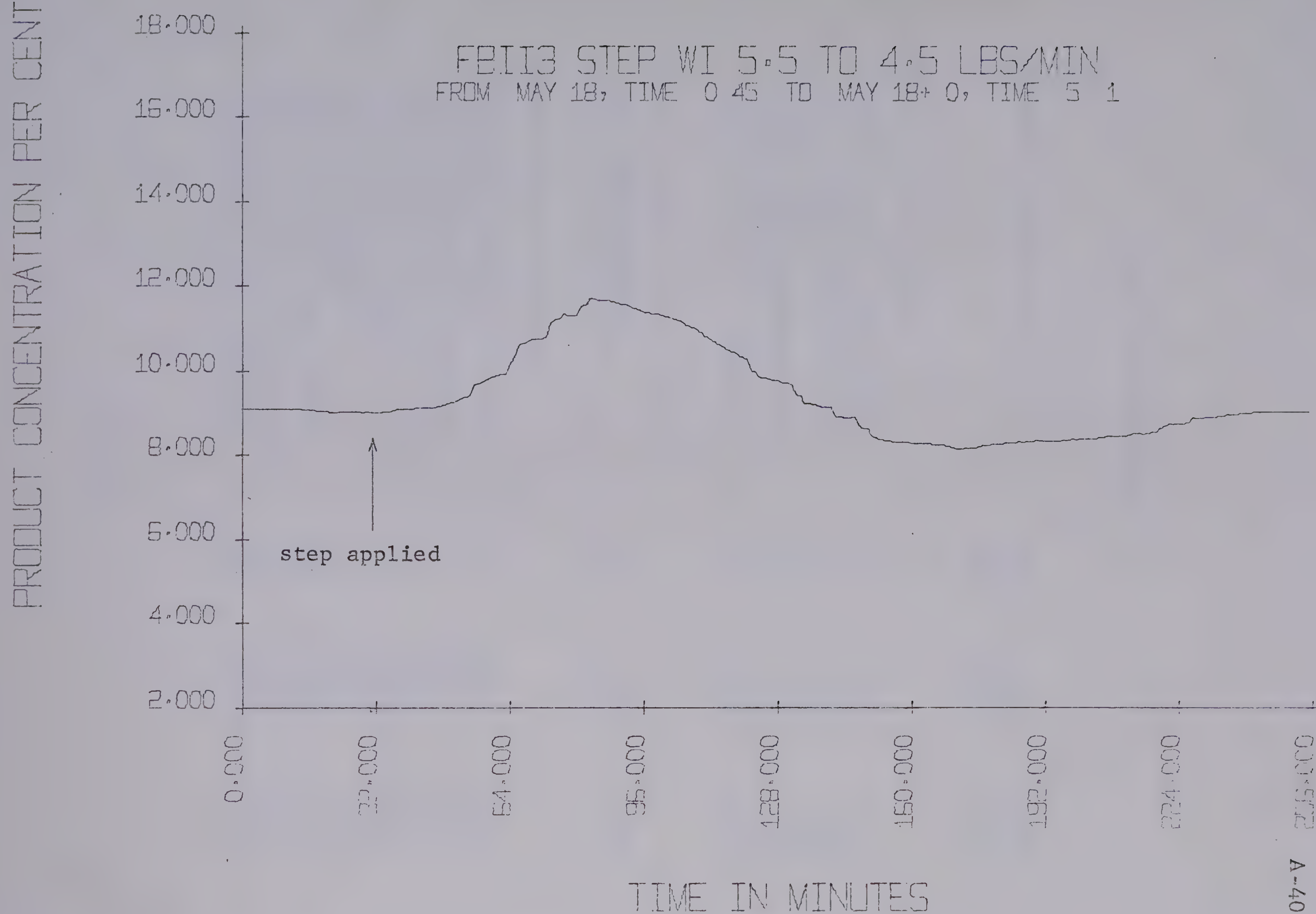

INFII 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 6 51 TO JUNE 7+ 0, TIME 8 59
PRODUCT CONCENTRATION PER CENT
18.000
16.000
14.000
12.000
10.000
8.000
6.000
4.000
2.000
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
step applied
process
+ + model
TIME IN MINUTES

Figure A-17

DATA FOR EXPERIMENT FB II 3

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.19	1.85
FIRST EFFECT BOTTOMS FLOW	F2	3.70	2.90
FIRST EFFECT OVERHEAD FLOW	F5	1.58	1.34
PRODUCT FLOW	F6	1.77	1.30
SECOND EFFECT OVERHEAD FLOW	F7	1.78	1.55
TOTAL FEED FLOW	F8	5.49	4.51
COOLING WATER TO CONDENSER	F9	49.40	49.24
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.80	2.80
PRODUCT CONCENTRATION	C6	9.04	9.06
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	101.70	96.80
FIRST EFFECT VAPOUR	T2	221.60	219.00
SOLUTION TO SECOND EFFECT	T4	181.40	181.20
STEAM CONDENSATE FIRST EFFECT	T5	241.40	235.90
FEED TO FIRST EFFECT	T7	190.30	189.20
STEAM TO SECOND EFFECT	T10	221.30	218.50
CONDENSER CONDENSATE	T11	121.80	111.60
SEPARATOR VAPOUR	T12	157.80	159.30
STEAM SUPPLY TO EVAPORATOR	T15	314.90	318.30
LIQUID IN FIRST EFFECT	T19	222.50	219.50
STEAM CONDENSATE SECOND EFFECT	T28	198.20	195.00
CONDENSER COOLING WATER IN	T29	65.00	65.00
PRODUCT FROM SECOND EFFECT	T34	195.90	196.60

TABLE A-20



FBII3 STEP WI 5.5 TO 4.5 LBS/MIN
FROM MAY 18, TIME 0 45 TO MAY 18+ 0, TIME 5 1
PRODUCT CONCENTRATION PER CENT
18.000
16.000
14.000
12.000
10.000
8.000
6.000
4.000
2.000
step applied
0.000
32.000
64.000
96.000
128.000
160.000
192.000
224.000
256.000
TIME IN MINUTES

Figure A-18

DATA FOR EXPERIMENT FF II 3

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.31	1.92
FIRST EFFECT BOTTOMS FLOW	F2	3.43	2.69
FIRST EFFECT OVERHEAD FLOW	F5	1.85	1.68
PRODUCT FLOW	F6	1.53	1.10
SECOND EFFECT OVERHEAD FLOW	F7	1.71	1.49
TOTAL FEED FLOW	F8	5.50	4.51
COOLING WATER TO CONDENSER	F9	43.38	50.06
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.00
PRODUCT CONCENTRATION	C6	10.28	10.73
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	111.00	105.80
FIRST EFFECT VAPOUR	T2	232.10	231.70
SOLUTION TO SECOND EFFECT	T4	191.30	190.40
STEAM CONDENSATE FIRST EFFECT	T5	248.40	244.70
FEED TO FIRST EFFECT	T7	196.30	195.30
STEAM TO SECOND EFFECT	T10	231.70	231.40
CONDENSER CONDENSATE	T11	138.60	98.60
SEPARATOR VAPOUR	T12	173.80	173.20
STEAM SUPPLY TO EVAPORATOR	T15	313.20	316.70
LIQUID IN FIRST EFFECT	T19	232.50	232.00
STEAM CONDENSATE SECOND EFFECT	T28	211.70	210.00
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	102.60	98.10

TABLE A-21

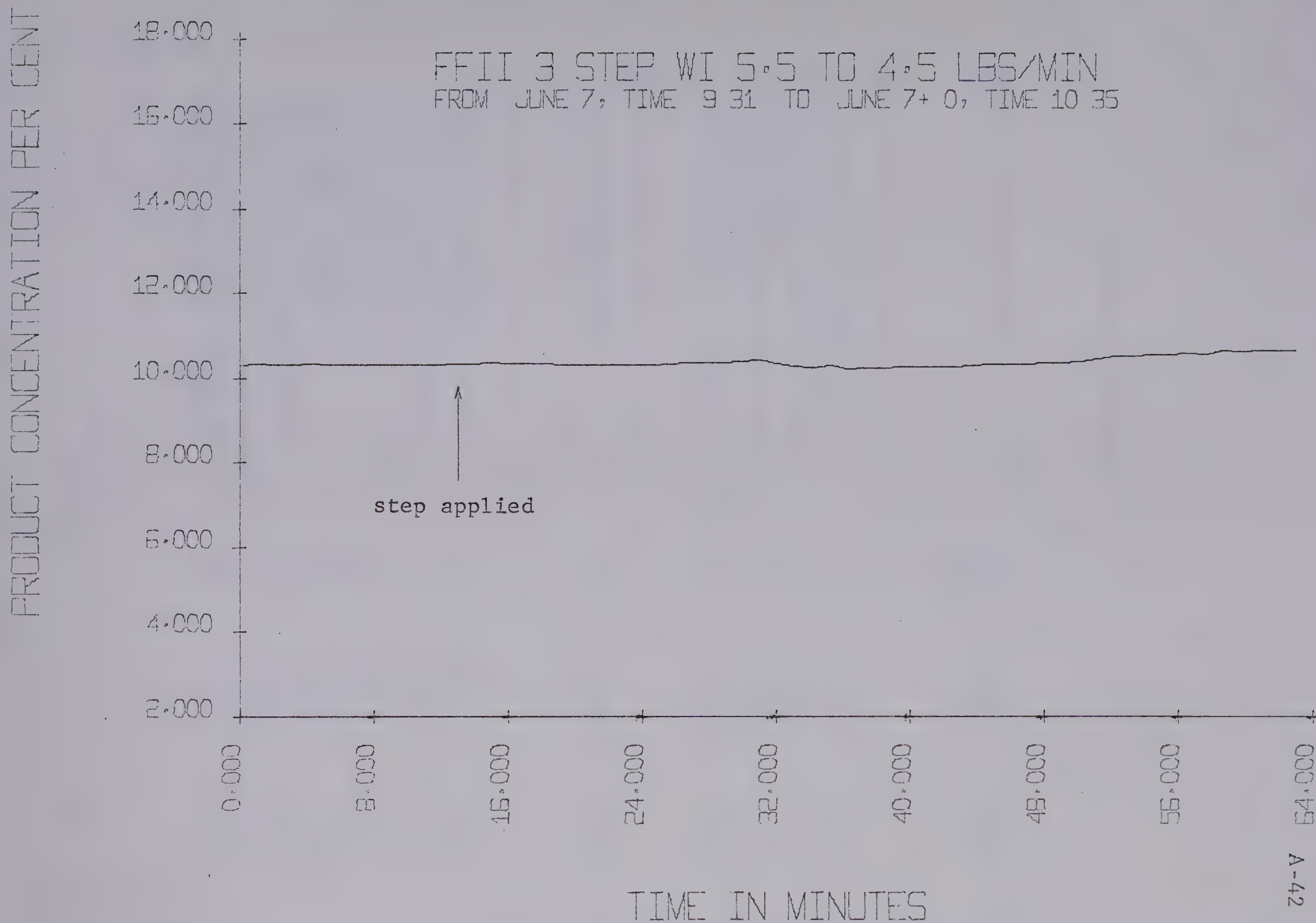

FFII 3 STEP WI 5.5 TO 4.5 LBS/MIN
FROM JUNE 7, TIME 9 31 TO JUNE 7+ 0, TIME 10 35
PRODUCT CONCENTRATION PER CENT
18.000
16.000
14.000
12.000
10.000
8.000
6.000
4.000
2.000
step applied
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
TIME IN MINUTES

Figure A-19

DATA FOR EXPERIMENT FF IV 1

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.31	1.80
FIRST EFFECT BOTTOMS FLOW	F2	3.52	2.79
FIRST EFFECT OVERHEAD FLOW	F5	2.01	1.57
PRODUCT FLOW	F6	1.52	1.19
SECOND EFFECT OVERHEAD FLOW	F7	1.63	1.37
TOTAL FEED FLOW	F8	5.51	4.48
COOLING WATER TO CONDENSER	F9	49.48	49.36
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.00
PRODUCT CONCENTRATION	C6	10.38	10.38
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	109.40	103.10
FIRST EFFECT VAPOUR	T2	231.70	228.00
SOLUTION TO SECOND EFFECT	T4	190.60	189.80
STEAM CONDENSATE FIRST EFFECT	T5	247.90	240.10
FEED TO FIRST EFFECT	T7	196.40	194.30
STEAM TO SECOND EFFECT	T10	231.30	227.40
CONDENSER CONDENSATE	T11	138.90	93.90
SEPARATOR VAPOUR	T12	172.70	172.10
STEAM SUPPLY TO EVAPORATOR	T15	312.90	317.70
LIQUID IN FIRST EFFECT	T19	232.60	228.30
STEAM CONDENSATE SECOND EFFECT	T28	210.90	206.90
CONDENSER COOLING WATER IN	T29	75.00	75.00
PRODUCT FROM SECOND EFFECT	T34	94.30	91.40

TABLE A-22

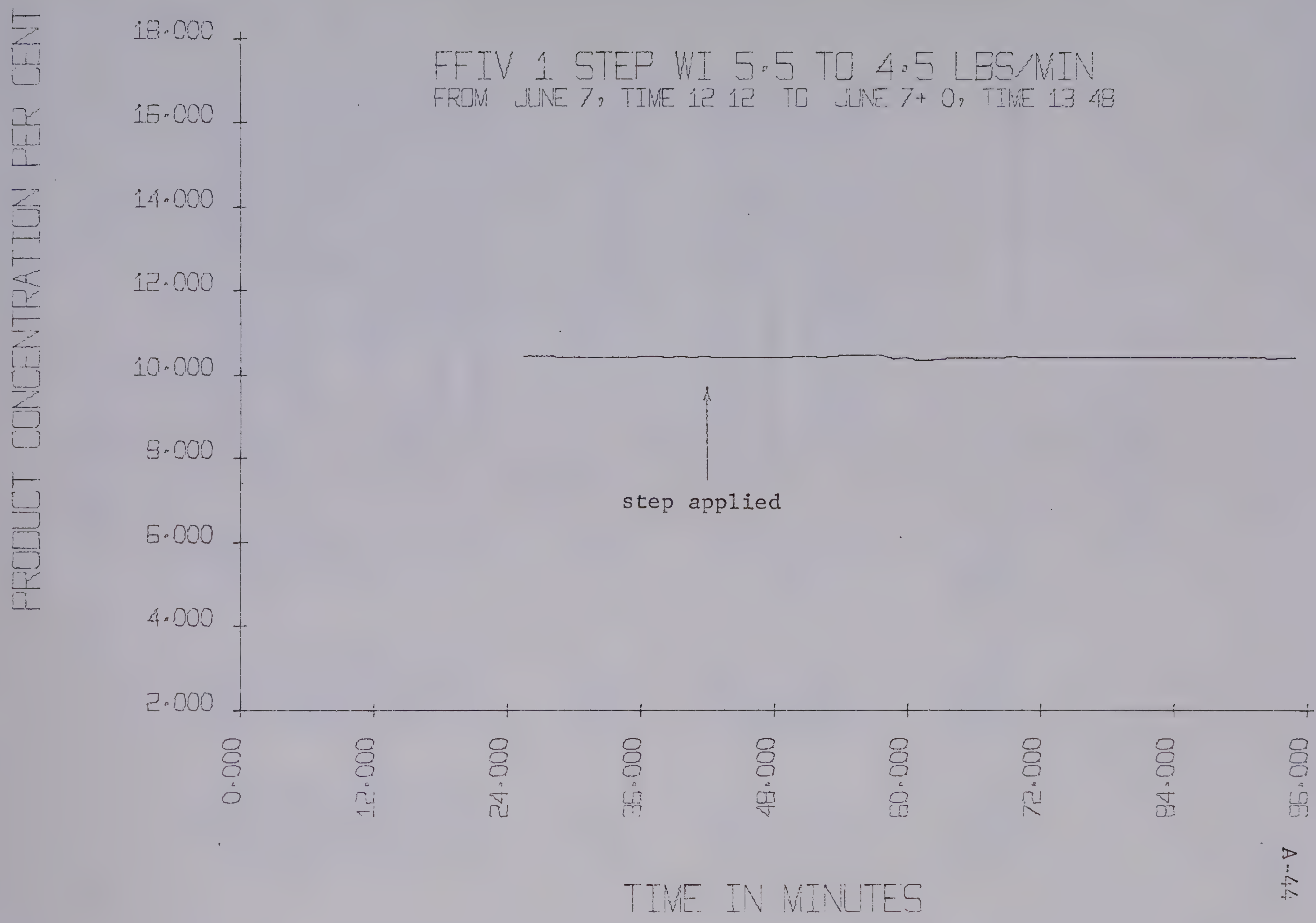

FFIV 1 STEP WI 5.5 TO 4.5 LBS/MIN
FROM JUNE 7, TIME 12 12 TO JUNE 7+ 0, TIME 13 48
PRODUCT CONCENTRATION PER CENT
18.000
16.000
14.000
12.000
10.000
8.000
6.000
4.000
2.000
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
step applied
TIME IN MINUTES

Figure A-20

DATA FOR EXPERIMENT FFB II 2

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.19	1.80
FIRST EFFECT BOTTOMS FLOW	F2	3.51	2.90
FIRST EFFECT OVERHEAD FLOW	F5	1.65	1.33
PRODUCT FLOW	F6	1.71	1.40
SECOND EFFECT OVERHEAD FLOW	F7	1.80	1.50
TOTAL FEED FLOW	F8	5.50	4.50
COOLING WATER TO CONDENSER	F9	48.78	49.18
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	2.80	2.80
PRODUCT CONCENTRATION	C6	8.59	8.53
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	102.80	97.30
FIRST EFFECT VAPOUR	T2	221.70	219.30
SOLUTION TO SECOND EFFECT	T4	182.80	180.90
STEAM CONDENSATE FIRST EFFECT	T5	240.90	235.00
FEED TO FIRST EFFECT	T7	189.60	188.30
STEAM TO SECOND EFFECT	T10	221.50	219.10
CONDENSER CONDENSATE	T11	122.80	113.20
SEPARATOR VAPOUR	T12	159.30	158.80
STEAM SUPPLY TO EVAPORATOR	T15	315.80	319.00
LIQUID IN FIRST EFFECT	T19	222.70	219.80
STEAM CONDENSATE SECOND EFFECT	T28	198.20	196.00
CONDENSER COOLING WATER IN	T29	65.00	65.00
PRODUCT FROM SECOND EFFECT	T34	197.80	197.50

TABLE A-23

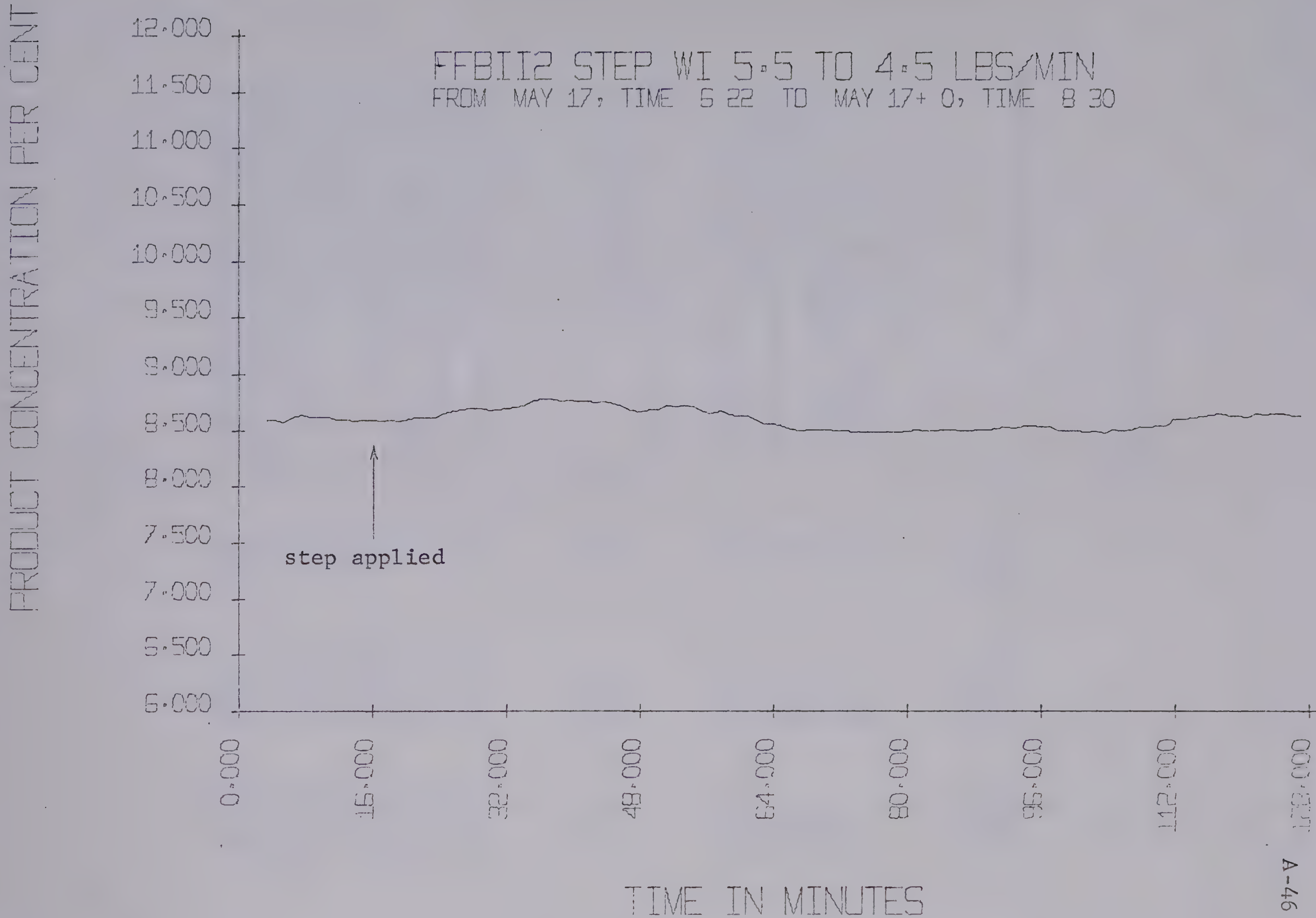
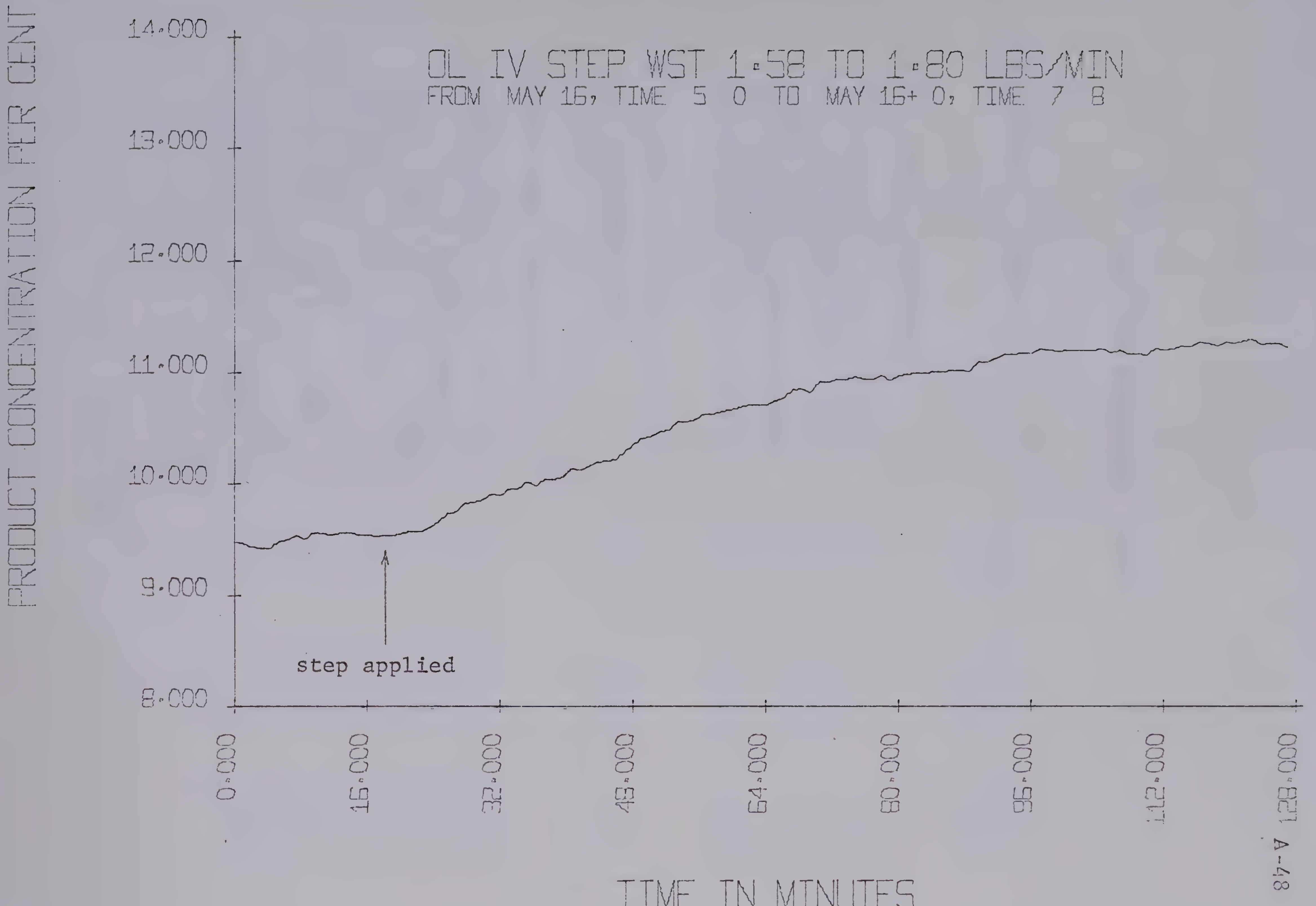

FFBII2 STEP WI 5.5 TO 4.5 LBS/MIN
FROM MAY 17, TIME 5 22 TO MAY 17+ 0, TIME 8 30
PRODUCT CONCENTRATION PER CENT
12.000
11.500
11.000
10.500
10.000
9.500
9.000
8.500
8.000
7.500
7.000
6.500
6.000
step applied
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
TIME IN MINUTES

Figure A-21

DATA FOR EXPERIMENT OL IV

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	1.58	1.80
FIRST EFFECT BOTTOMS FLOW	F2	3.01	2.78
FIRST EFFECT OVERHEAD FLOW	F5	1.30	1.52
PRODUCT FLOW	F6	1.62	1.27
SECOND EFFECT OVERHEAD FLOW	F7	1.27	1.40
TOTAL FEED FLOW	F8	4.53	4.48
COOLING WATER TO CONDENSER	F9	49.49	48.79
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.18	3.07
PRODUCT CONCENTRATION	C6	9.56	11.29
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	95.20	97.40
FIRST EFFECT VAPOUR	T2	210.10	219.80
SOLUTION TO SECOND EFFECT	T4	177.20	177.20
STEAM CONDENSATE FIRST EFFECT	T5	226.10	237.90
FEED TO FIRST EFFECT	T7	191.00	191.30
STEAM TO SECOND EFFECT	T10	209.90	219.50
CONDENSER CONDENSATE	T11	110.10	117.30
SEPARATOR VAPOUR	T12	152.80	152.90
STEAM SUPPLY TO EVAPORATOR	T15	318.60	317.10
LIQUID IN FIRST EFFECT	T19	211.10	220.90
STEAM CONDENSATE SECOND EFFECT	T28	194.20	192.30
CONDENSER COOLING WATER IN	T29	65.00	65.00
PRODUCT FROM SECOND EFFECT	T34	192.70	194.20

TABLE A-24



OL IV STEP WST 1.58 TO 1.80 LBS/MIN
FROM MAY 16, TIME 5 0 TO MAY 16+ 0, TIME 7 8
PRODUCT CONCENTRATION PER CENT
14.000
13.000
12.000
11.000
10.000
9.000
8.000
step applied
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
TIME IN MINUTES

Figure A-22

DATA FOR EXPERIMENT CONINF

VARIABLE DESCRIPTION	VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS/MIN			
STEAM FLOW TO FIRST EFFECT	F1	2.01	1.78
FIRST EFFECT BOTTOMS FLOW	F2	3.65	3.32
FIRST EFFECT OVERHEAD FLOW	F5	1.70	1.46
PRODUCT FLOW	F6	2.03	1.86
SECOND EFFECT OVERHEAD FLOW	F7	1.55	1.35
TOTAL FEED FLOW	F8	5.51	5.01
COOLING WATER TO CONDENSER	F9	42.89	42.99
CONCENTRATIONS WEIGHT PER CENT			
FEED CONCENTRATION	C1	3.00	3.59
PRODUCT CONCENTRATION	C6	7.81	9.02
TEMPERATURES IN DEGREES F			
CONDENSER COOLING WATER OUT	T1	104.10	101.30
FIRST EFFECT VAPOUR	T2	222.40	223.50
SOLUTION TO SECOND EFFECT	T4	184.80	185.60
STEAM CONDENSATE FIRST EFFECT	T5	240.80	240.00
FEED TO FIRST EFFECT	T7	191.10	190.40
STEAM TO SECOND EFFECT	T10	222.10	223.20
CONDENSER CONDENSATE	T11	117.10	91.60
SEPARATOR VAPOUR	T12	164.50	166.00
STEAM SUPPLY TO EVAPORATOR	T15	312.40	314.50
LIQUID IN FIRST EFFECT	T19	222.80	224.00
STEAM CONDENSATE SECOND EFFECT	T28	197.30	196.90
CONDENSER COOLING WATER IN	T29	70.00	70.00
PRODUCT FROM SECOND EFFECT	T34	209.40	210.80

TABLE A-25

The following six pages contain graphs showing the transient behaviour of the steam flow for all experiments in which a positive step in feed flow rate was applied. They have been included here to illustrate the controller output obtained for different types of controllers.

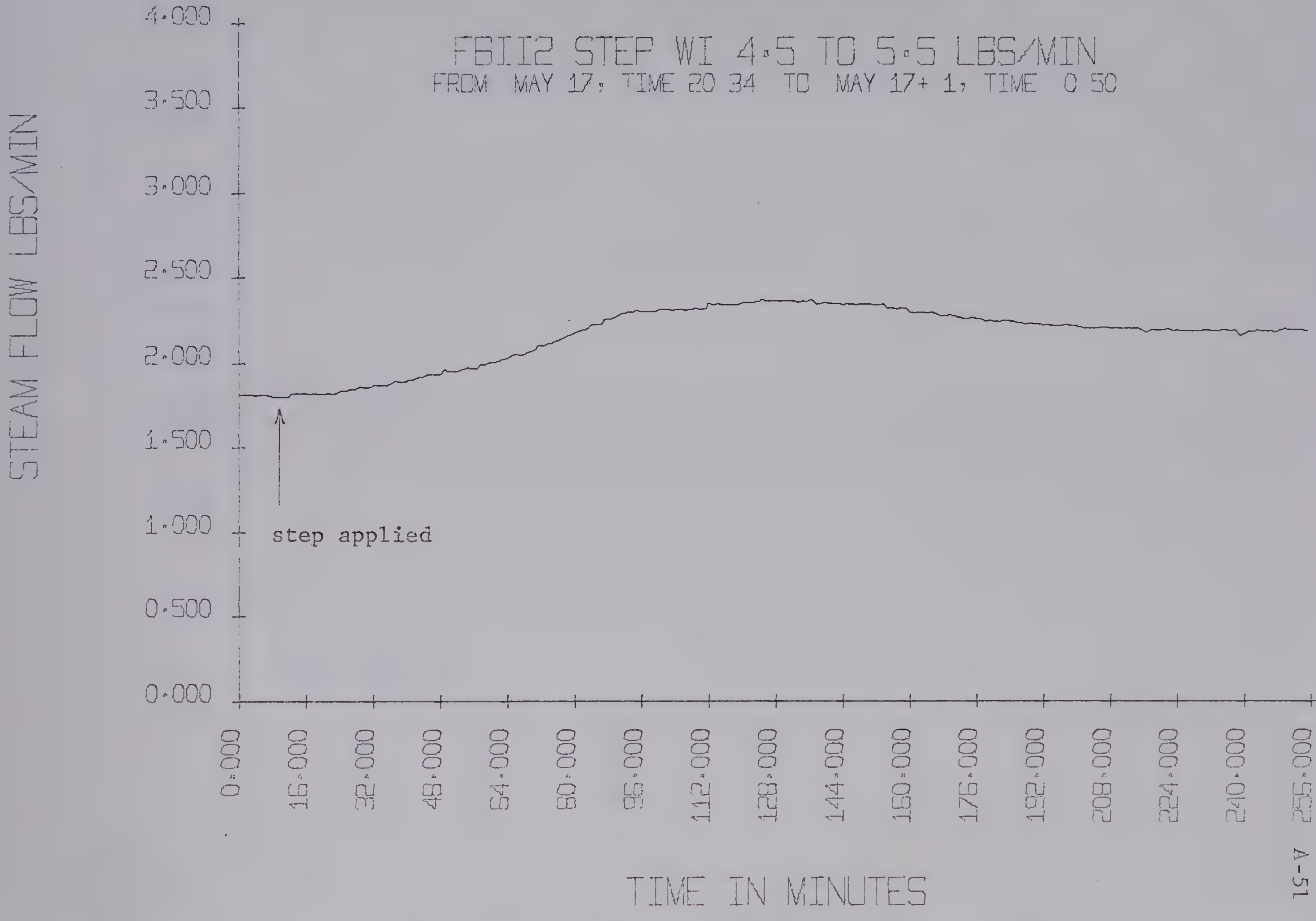
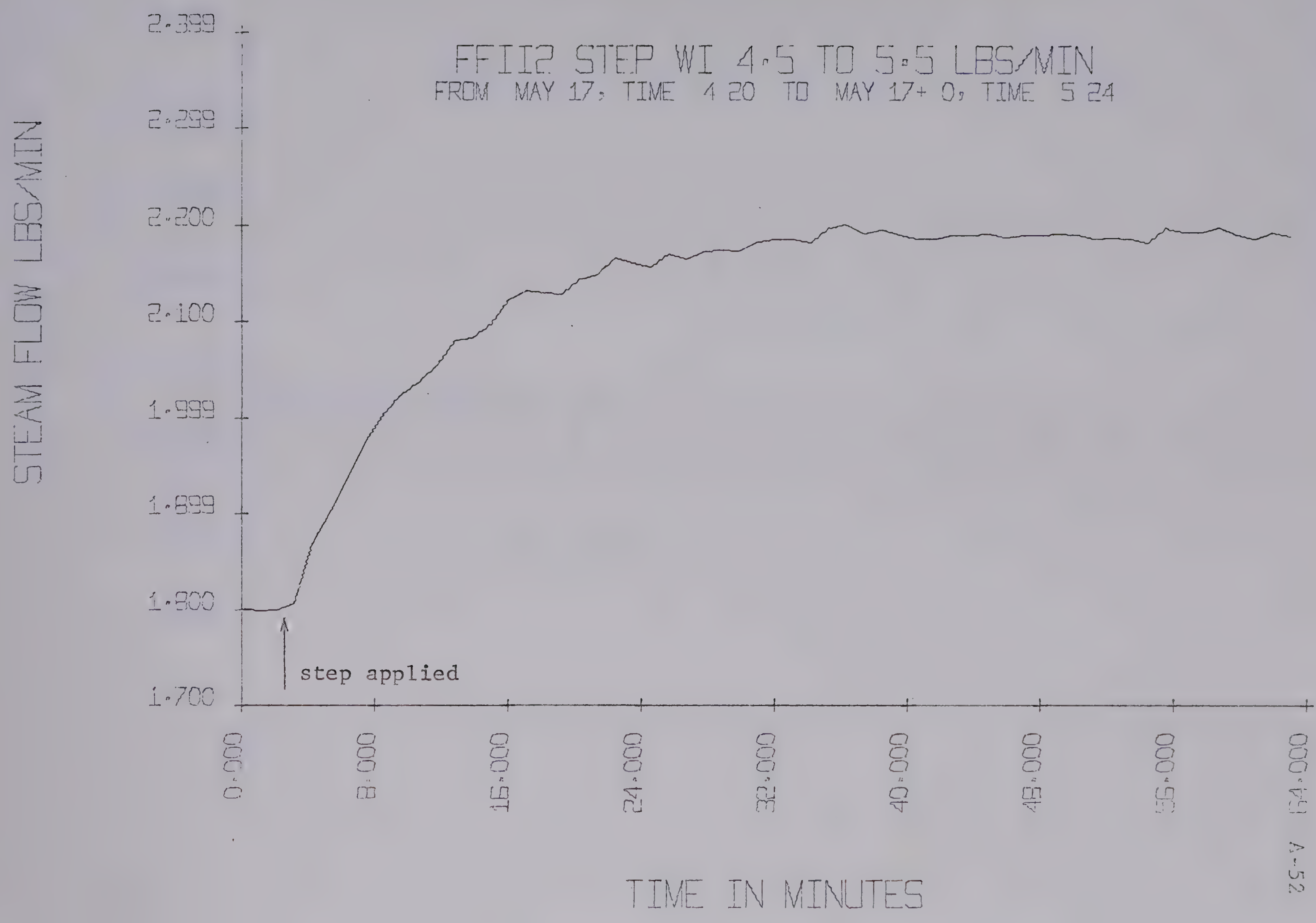

FBII2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM MAY 17, TIME 20 34 TO MAY 17+ 1, TIME 0 50
STEAM FLOW LBS/MIN
4.000
3.500
3.000
2.500
2.000
1.500
1.000
0.500
0.000
step applied
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
144.000
160.000
176.000
192.000
208.000
224.000
240.000
256.000
TIME IN MINUTES

Figure A-23



FFII2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM MAY 17, TIME 4 20 TO MAY 17+ 0, TIME 5 24
STEAM FLOW LBS/MIN
2.399
2.299
2.200
2.100
1.999
1.899
1.800
1.700
step applied
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
TIME IN MINUTES

Figure A-24

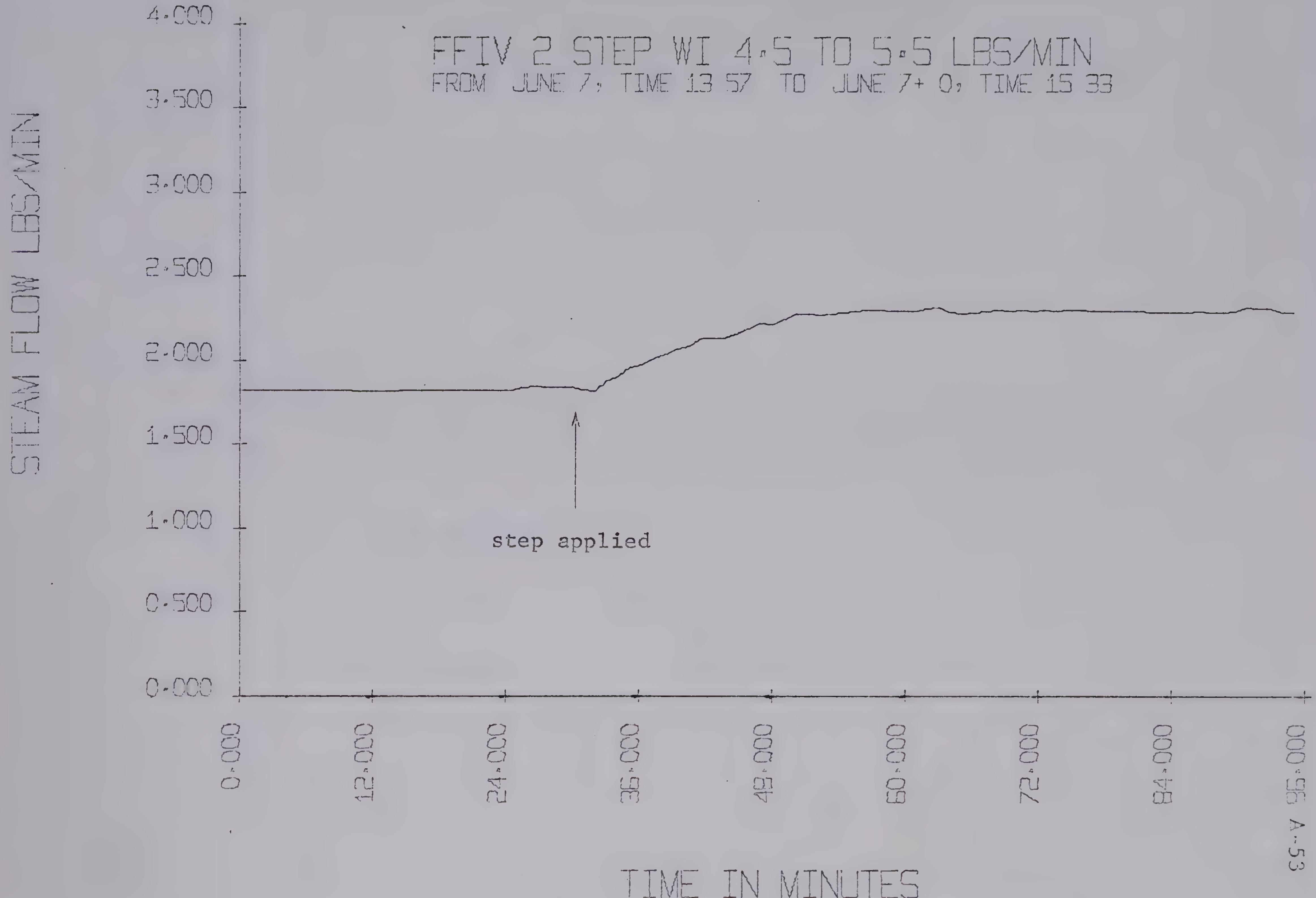

FFIV 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 13 57 TO JUNE 7+ 0, TIME 15 33
STEAM FLOW LBS/MIN
4.000
3.500
3.000
2.500
2.000
1.500
1.000
0.500
0.000
step applied
0.000
12.000
24.000
36.000
48.000
60.000
72.000
84.000
96.000
TIME IN MINUTES

Figure A-25

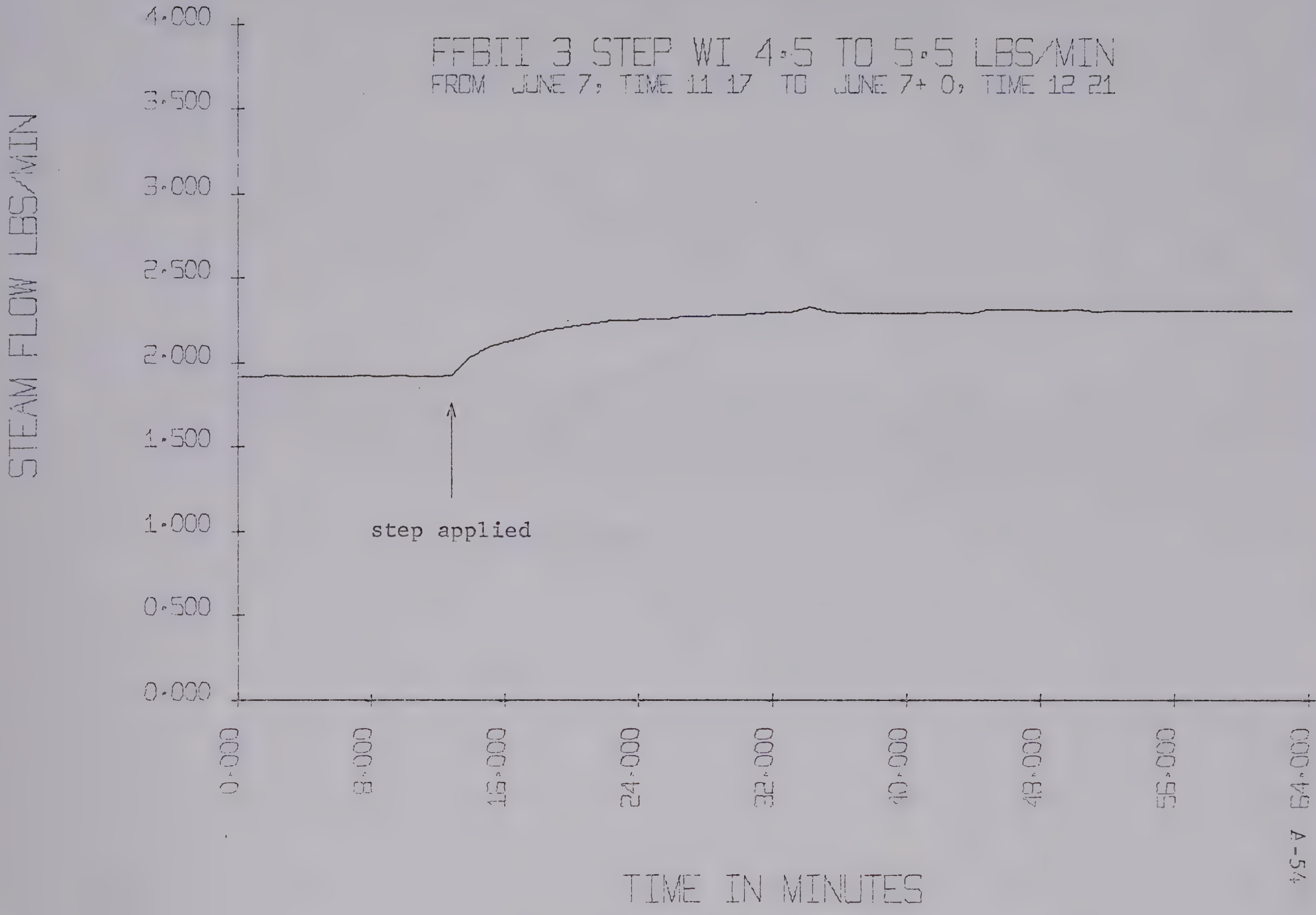

FFBII 3 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 11 17 TO JUNE 7+ 0, TIME 12 21
STEAM FLOW LBS/MIN
4.000
3.500
3.000
2.500
2.000
1.500
1.000
0.500
0.000
step applied
0.000
8.000
16.000
24.000
32.000
40.000
48.000
56.000
64.000
TIME IN MINUTES

Figure A-26

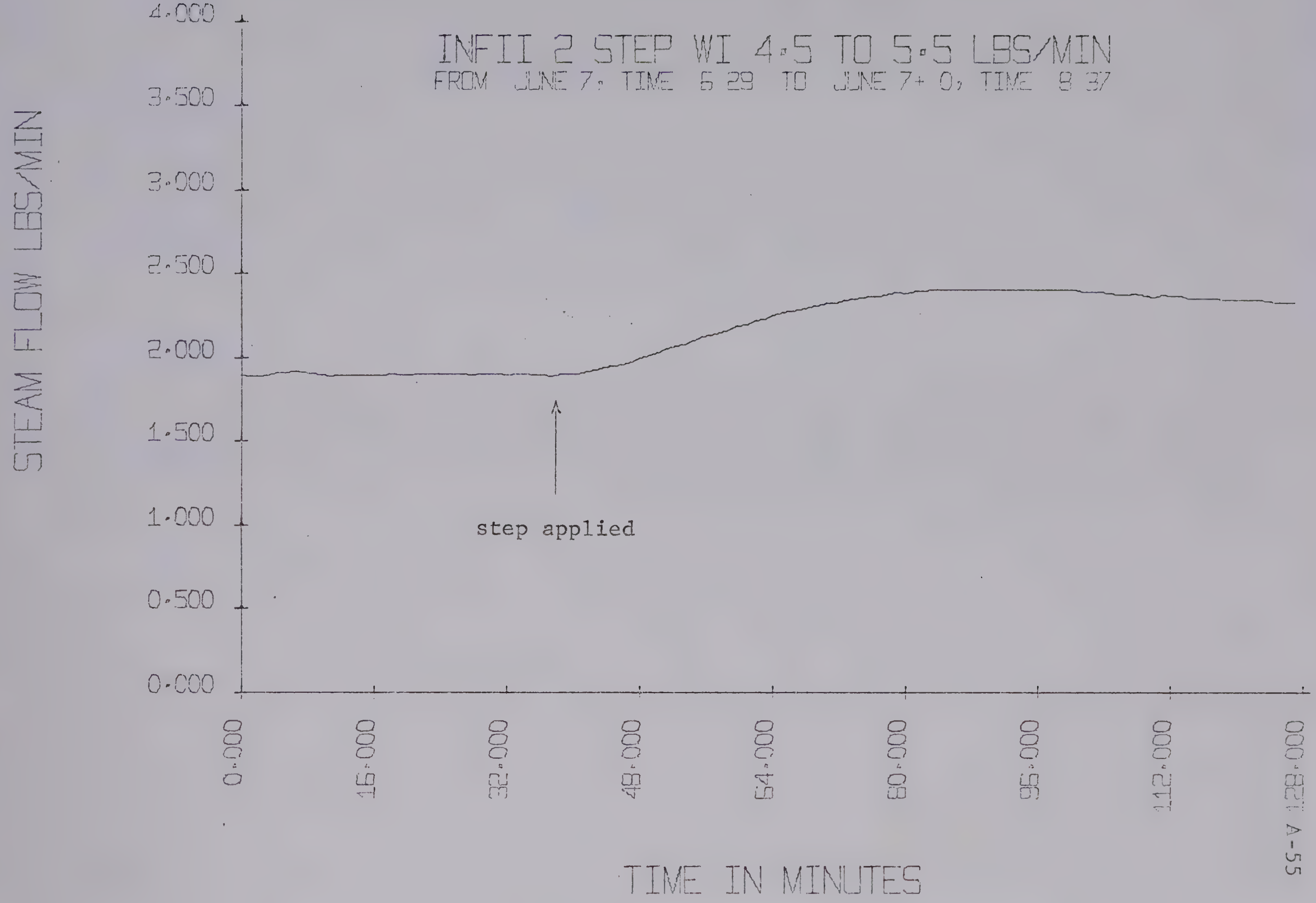

INFII 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 6 29 TO JUNE 7+ 0, TIME 8 37
STEAM FLOW LBS/MIN
4.000
3.500
3.000
2.500
2.000
1.500
1.000
0.500
0.000
step applied
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
TIME IN MINUTES

Figure A-27

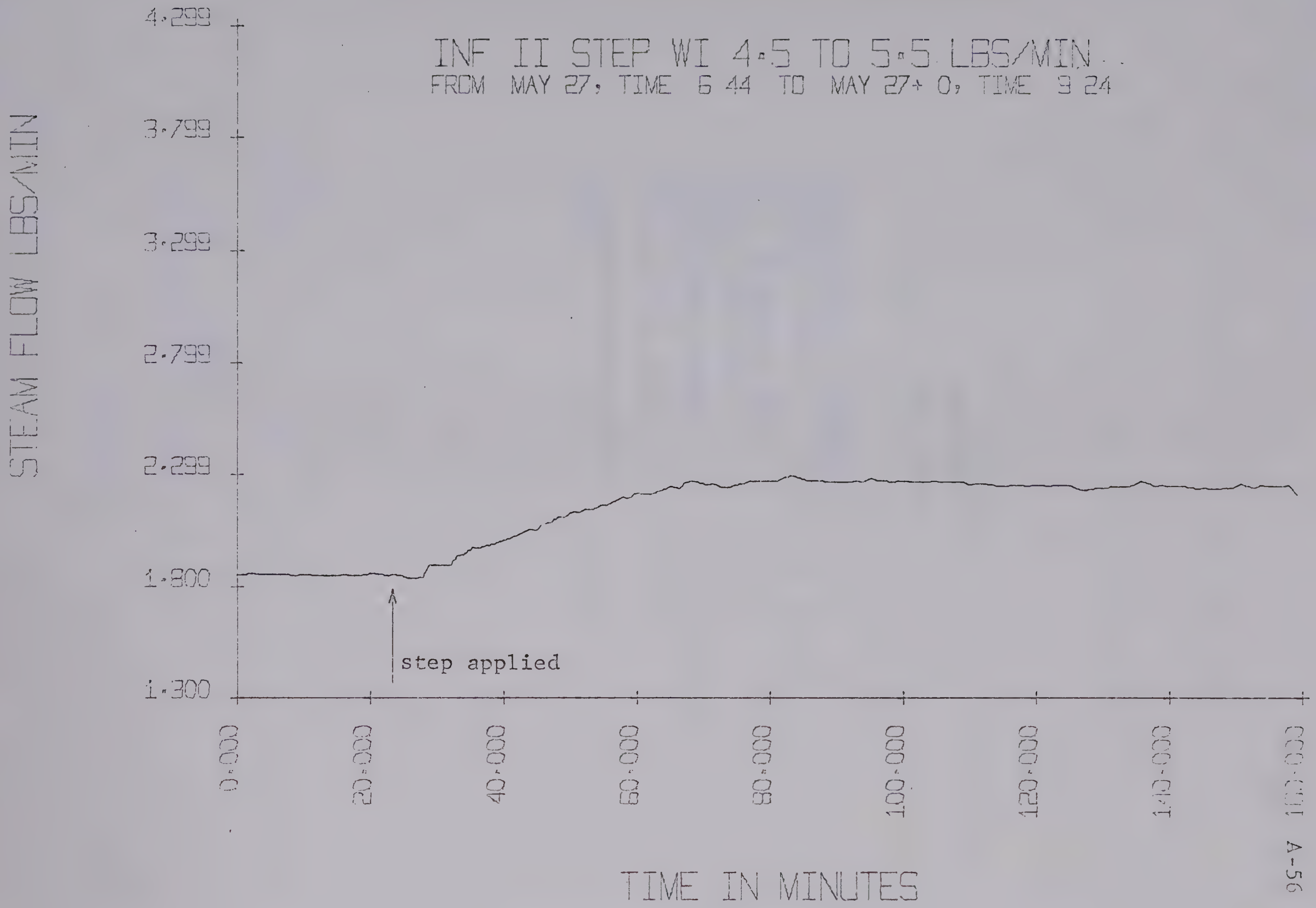

INF II STEP WI 4.5 TO 5.5 LBS/MIN
FROM MAY 27, TIME 6 44 TO MAY 27+ 0, TIME 9 24
STEAM FLOW LBS/MIN
4.299
3.799
3.299
2.799
2.299
1.800
1.300
step applied
0.000
20.000
40.000
60.000
80.000
100.000
120.000
140.000
TIME IN MINUTES

Figure A-23

Appendix B

Determination of Transportation Lags

Feed Piping

pipe diameter 3/8 in i.d.

pipe length 27 ft

pipe cross section 0.11 in^2

equivalent lengths of resistances (25)

2 globe valves 16 ft each
7 elbows 3 ft each
orifice and refractometer 0.6 ft each

total equivalent pipe length 81 ft

flow rate 5.0 lbs/min specific gravity 1.0

velocity 105 ft/min

transportation lag 81/1o5 = 0.78 min

Interstage Piping

pipe diameter 3/8 in i.d.

pipe length 11 ft

pipe cross section 0.11 in^2

equivalent lengths of resistances

3 globe valves 16 ft each
1 orifice 0.6 ft

total equivalent pipe length 60 ft

flow rate 3.30 lbs/min specific gravity 1.0

velocity 69 ft/min

transportation lag 60/69 = 0.86 min

Table B- 1

List of Orifice and Valve Stem Sizes

Variable Concerned	Orifice Diameter Inches	Valve Constant C_V
F1	0.382	2.25
F2	0.125	1.50
F5	0.100	0.56
F6	0.100	0.56
F7	0.150	
F8	0.160	
F9	0.327	2.25
F10	0.875	
F11	0.125	0.20
F12	0.075	0.125
L12		1.13
L13		0.56
DVP21		0.012
P22		0.012
TT10		0.28
TT11		0.28

Table B- 2

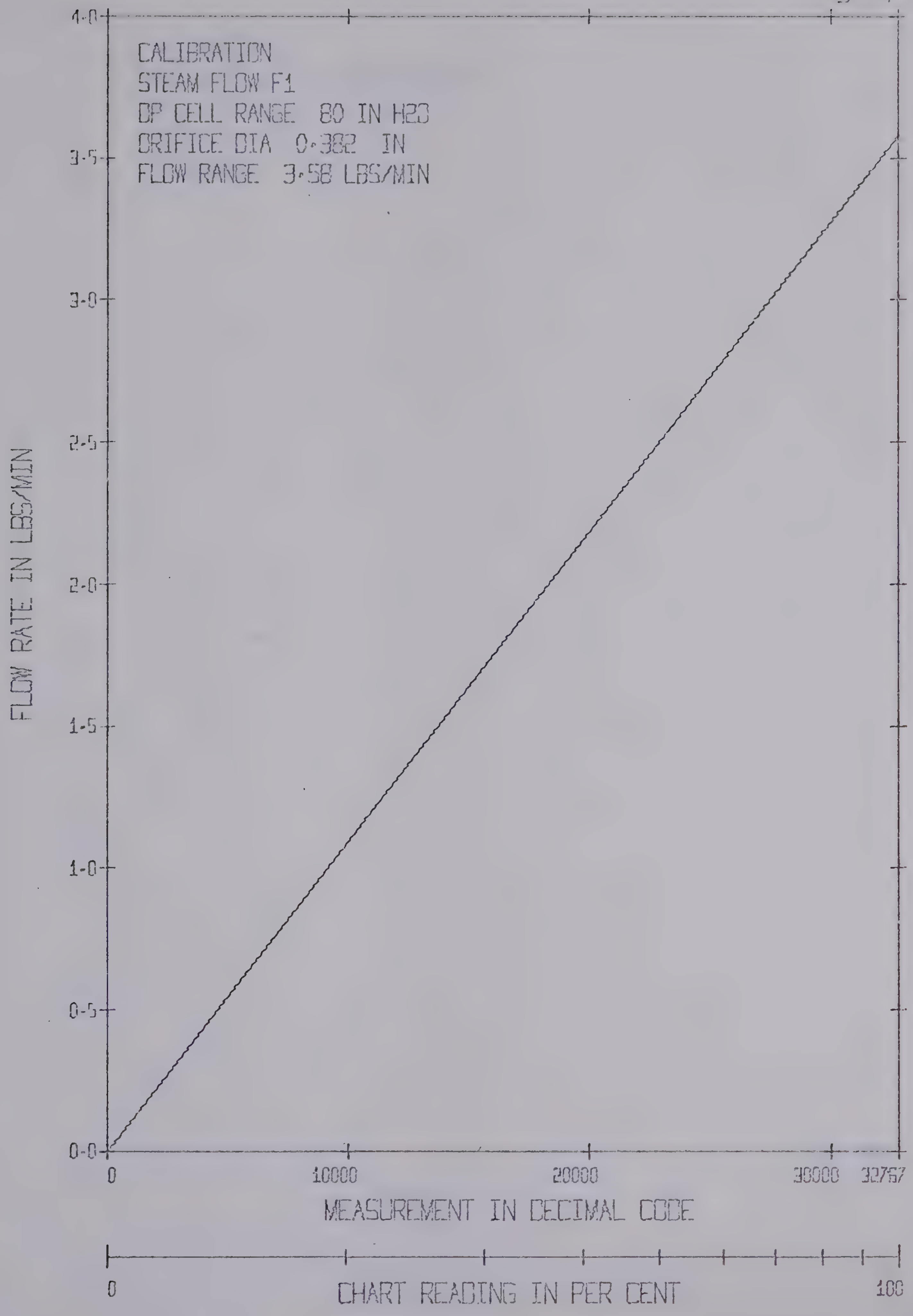

CALIBRATION
STEAM FLOW F1
DP CELL RANGE 80 IN H2O
ORIFICE DIA 0.382 IN
FLOW RANGE 3.58 LBS/MIN
FLOW RATE IN LBS/MIN
0.0
0.5
1.0
1.5
2.0
2.5
3.0
3.5
4.0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-1

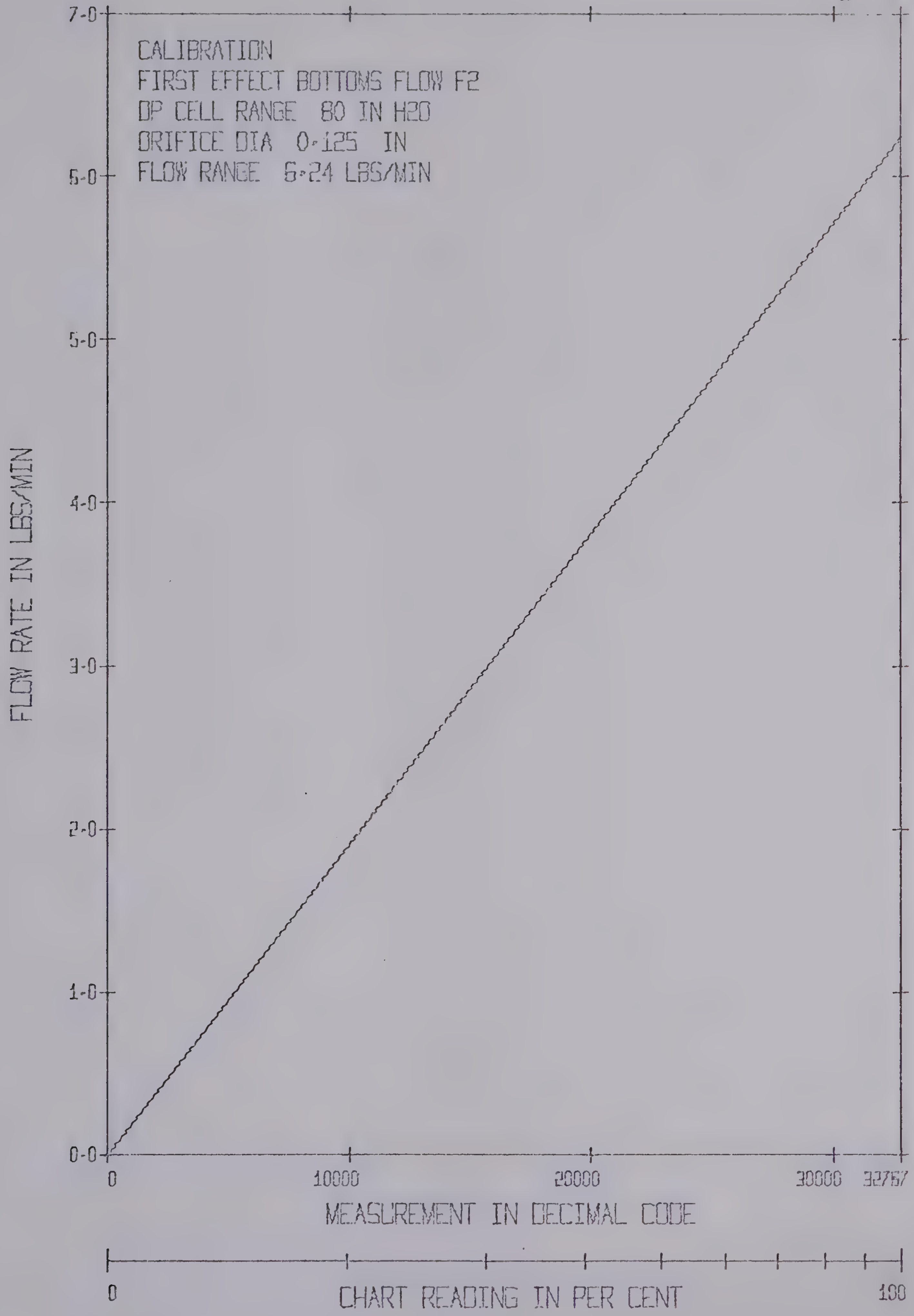

CALIBRATION
FIRST EFFECT BOTTOMS FLOW F2
DP CELL RANGE 80 IN H2O
ORIFICE DIA 0.125 IN
FLOW RANGE 6.24 LBS/MIN
FLOW RATE IN LBS/MIN
7.0
6.0
5.0
4.0
3.0
2.0
1.0
0.0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-2

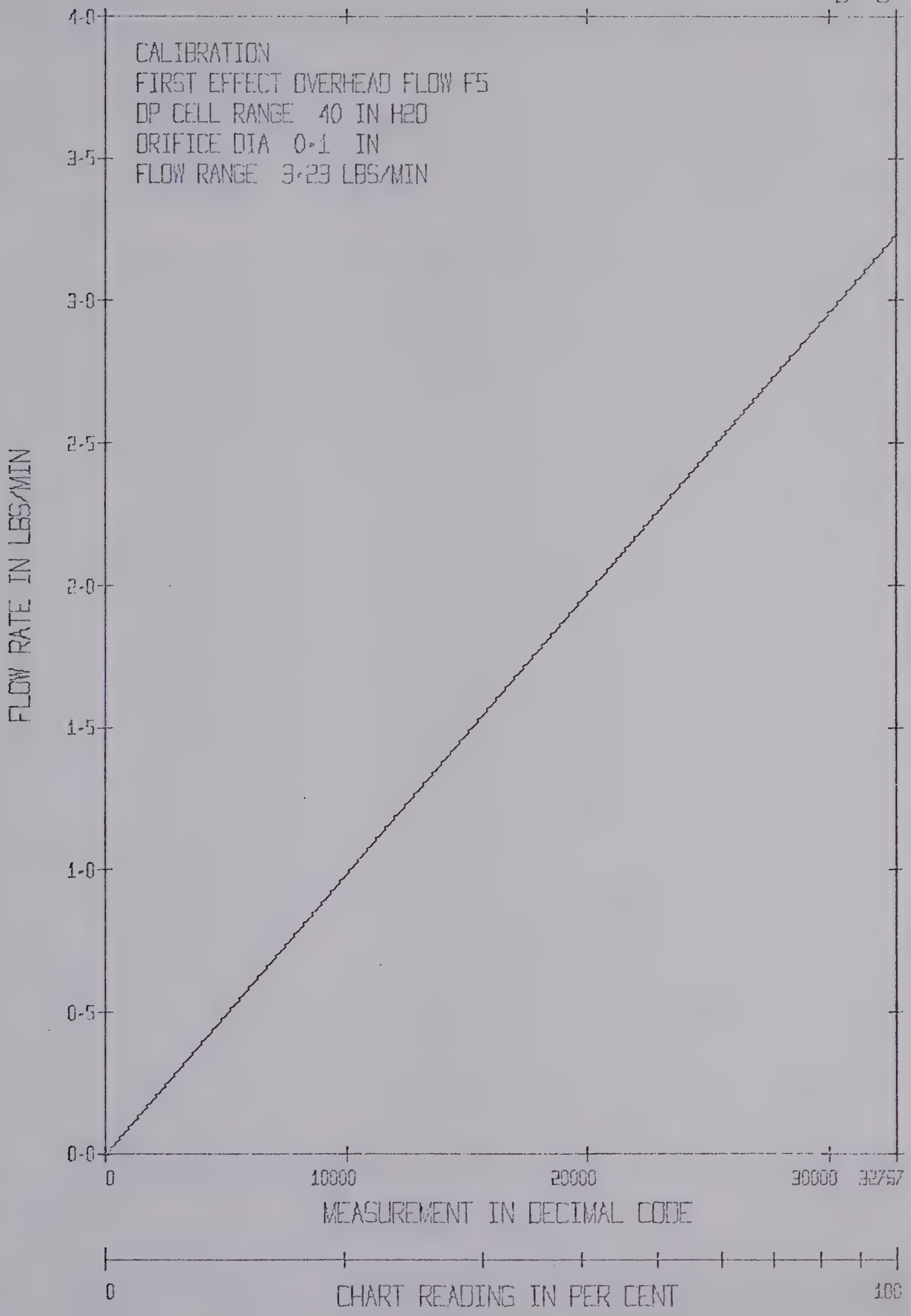

CALIBRATION
FIRST EFFECT OVERHEAD FLOW F5
DP CELL RANGE 40 IN H2O
ORIFICE DIA 0.1 IN
FLOW RANGE 3.23 LBS/MIN
FLOW RATE IN LBS/MIN
4.0
3.5
3.0
2.5
2.0
1.5
1.0
0.5
0.0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-3

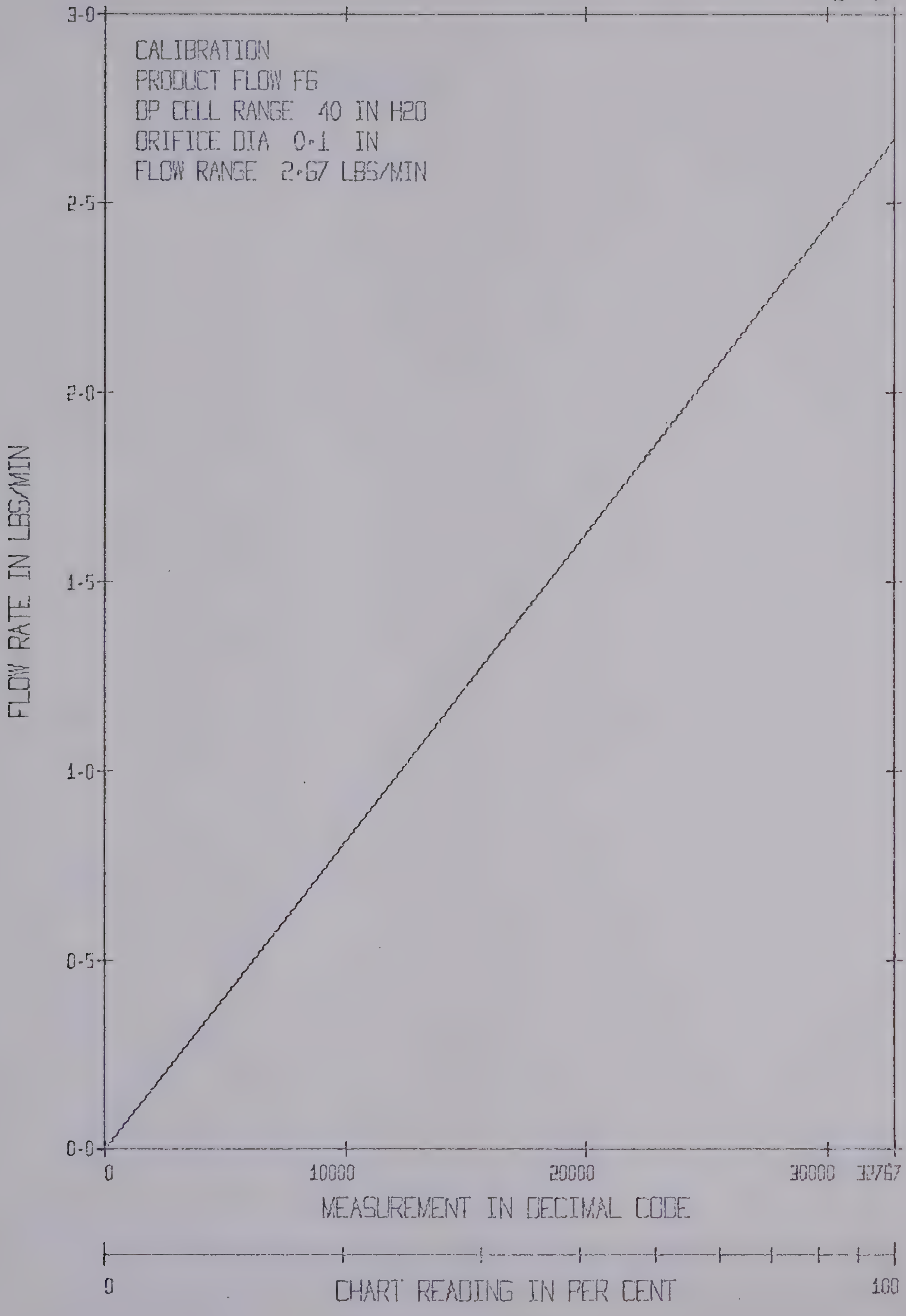

CALIBRATION
PRODUCT FLOW F6
DP CELL RANGE 40 IN H2O
ORIFICE DIA 0·1 IN
FLOW RANGE 2·67 LBS/MIN
3·0
2·5
2·0
1·5
1·0
0·5
0·0
FLOW RATE IN LBS/MIN
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-4

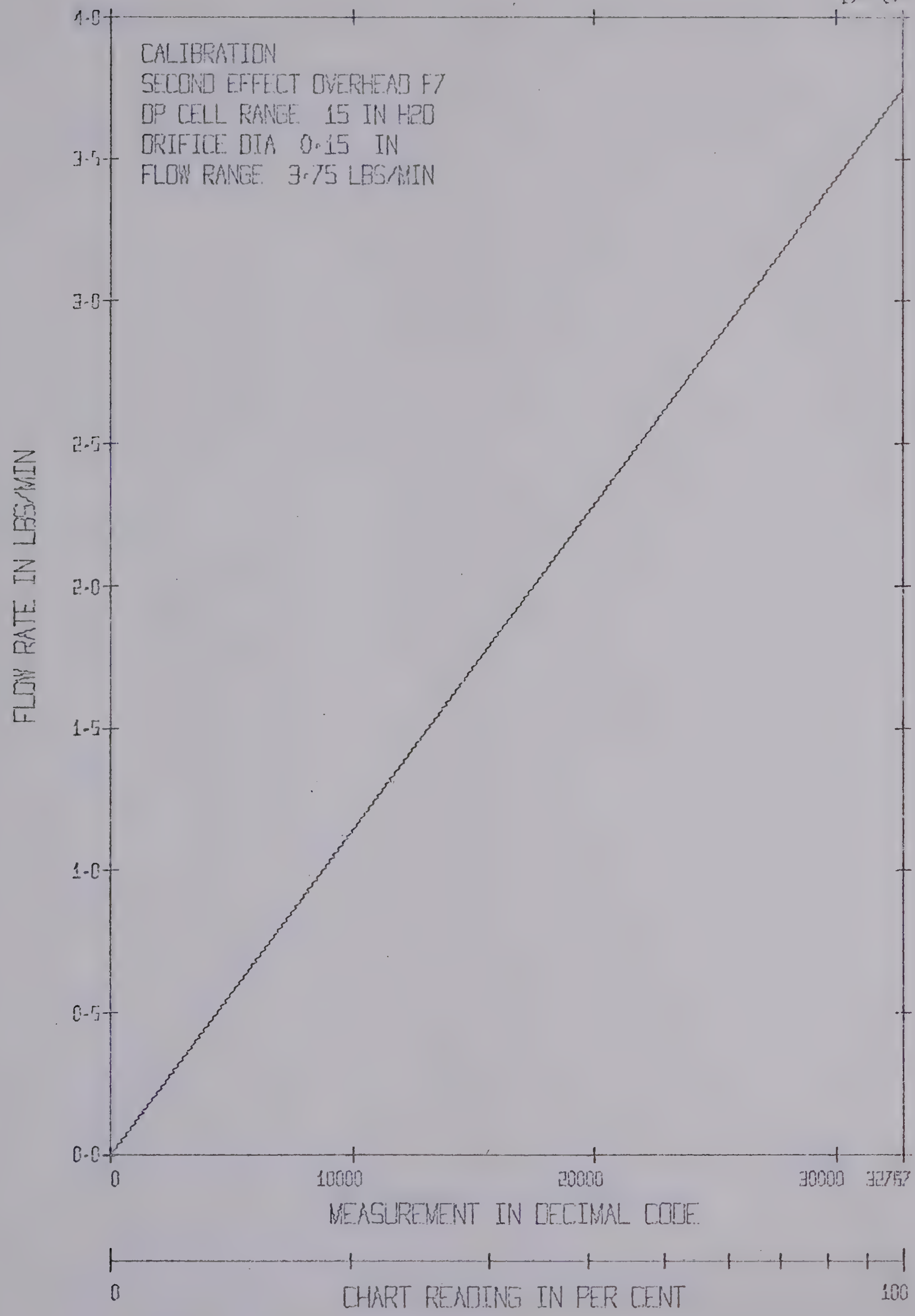

CALIBRATION
SECOND EFFECT OVERHEAD F7
DP CELL RANGE 15 IN H2O
ORIFICE DIA 0.15 IN
FLOW RANGE 3.75 LBS/MIN
FLOW RATE IN LBS/MIN
0.0
0.5
1.0
1.5
2.0
2.5
3.0
3.5
4.0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-5

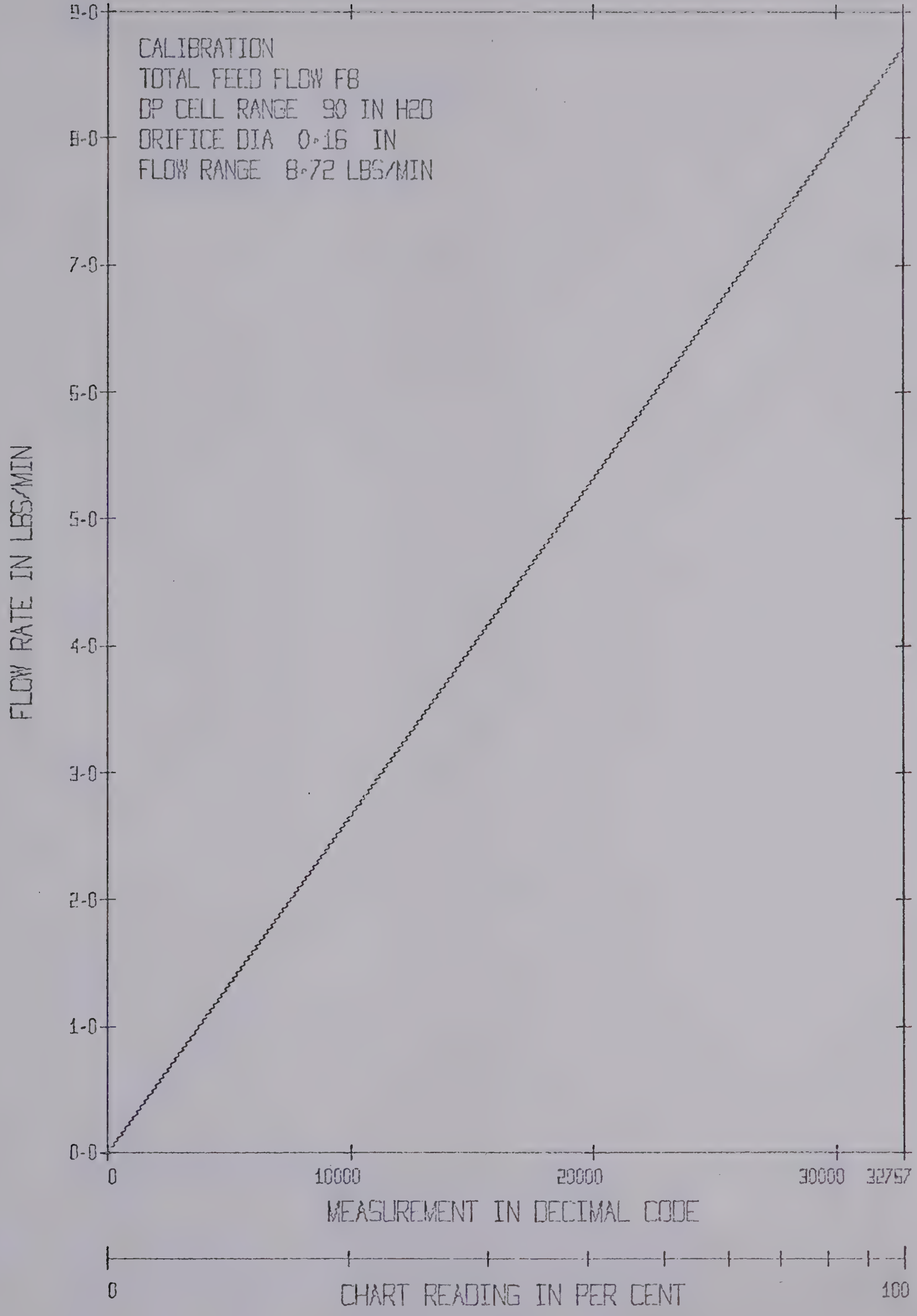

CALIBRATION
TOTAL FEED FLOW F8
DP CELL RANGE 90 IN H2O
ORIFICE DIA 0.16 IN
FLOW RANGE 8.72 LBS/MIN
FLOW RATE IN LBS/MIN
0.0
1.0
2.0
3.0
4.0
5.0
6.0
7.0
8.0
9.0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-6

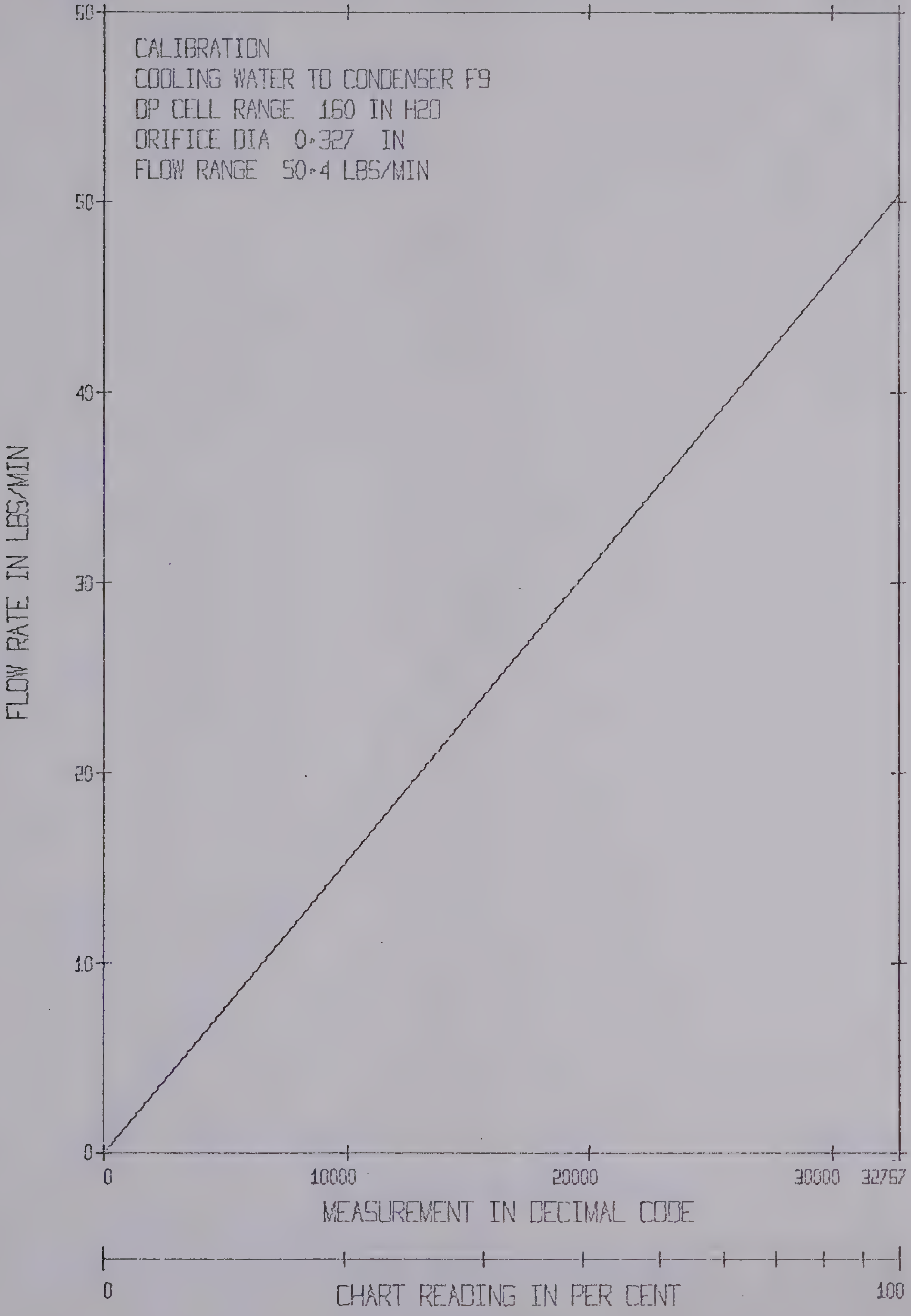

CALIBRATION
COOLING WATER TO CONDENSER F9
DP CELL RANGE 160 IN H2O
ORIFICE DIA 0.327 IN
FLOW RANGE 50.4 LBS/MIN
FLOW RATE IN LBS/MIN
0
10
20
30
40
50
60
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
100
CHART READING IN PER CENT

FIGURE B-7

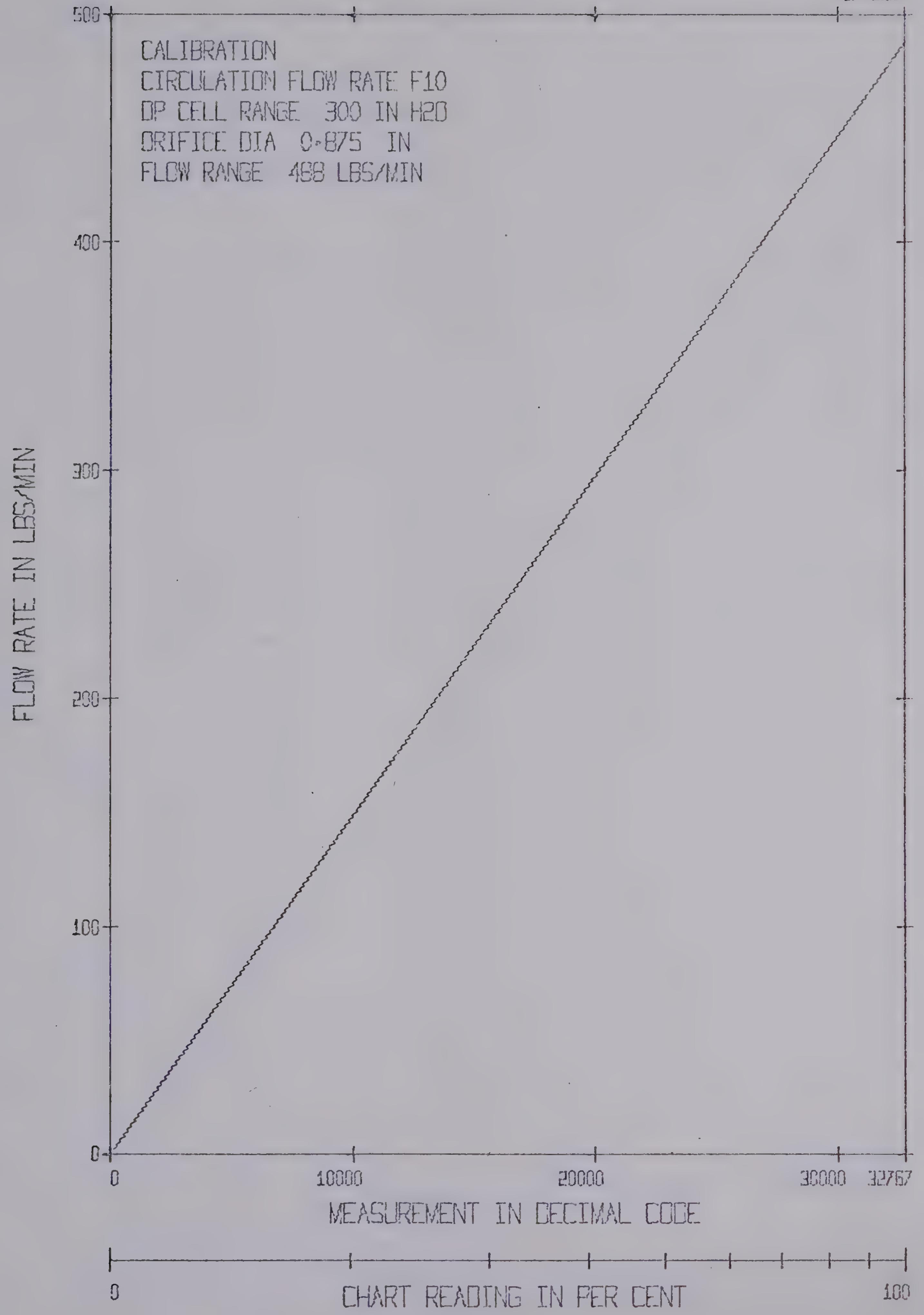

CALIBRATION
CIRCULATION FLOW RATE F10
DP CELL RANGE 300 IN H2O
ORIFICE DIA 0.875 IN
FLOW RANGE 488 LBS/MIN
FLOW RATE IN LBS/MIN
0
100
200
300
400
500
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
100
CHART READING IN PER CENT

FIGURE B-8

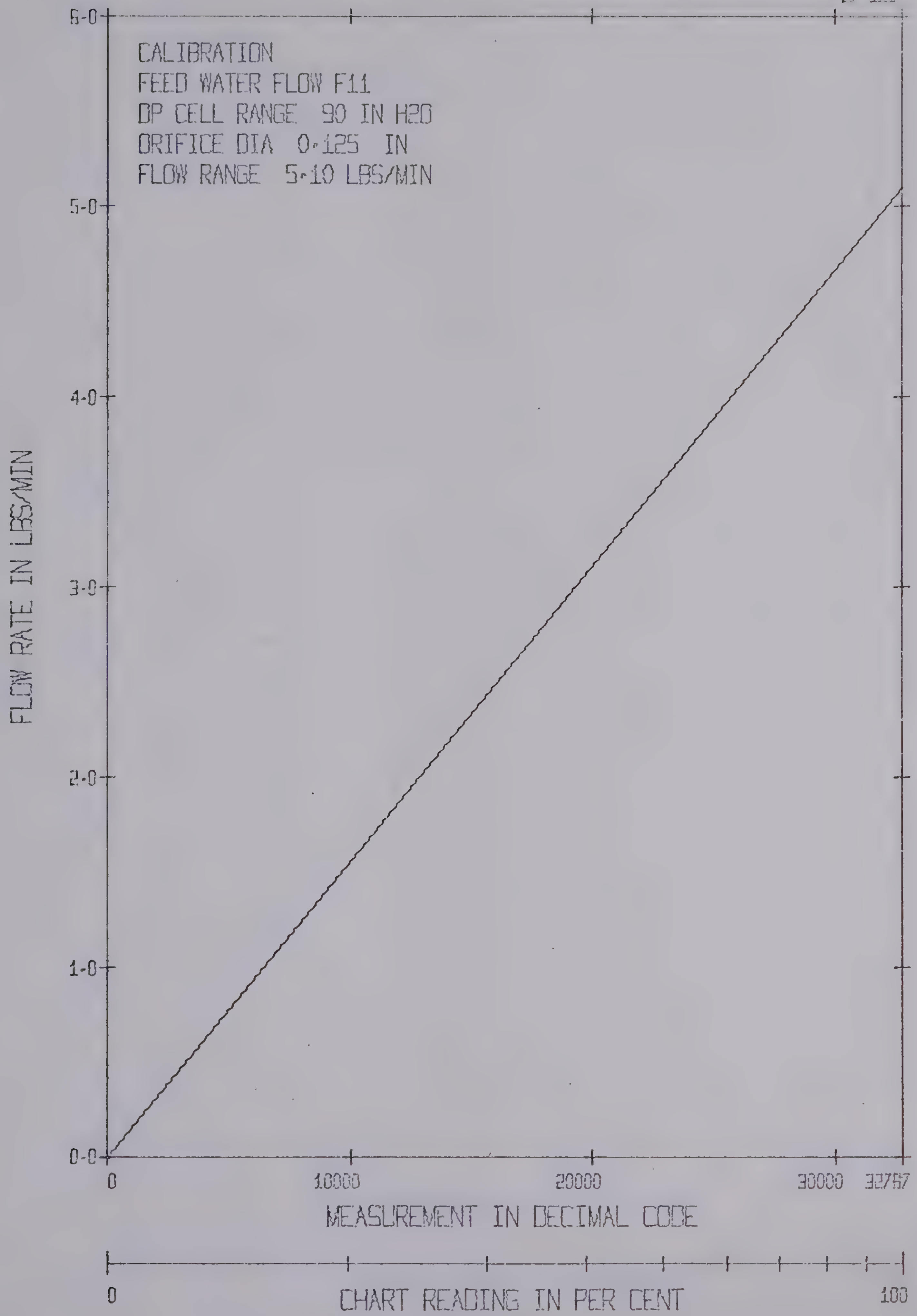

CALIBRATION
FEED WATER FLOW F11
DP CELL RANGE 90 IN H2O
ORIFICE DIA 0.125 IN
FLOW RANGE 5-10 LBS/MIN
FLOW RATE IN LBS/MIN
6.0
5.0
4.0
3.0
2.0
1.0
0.0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-9

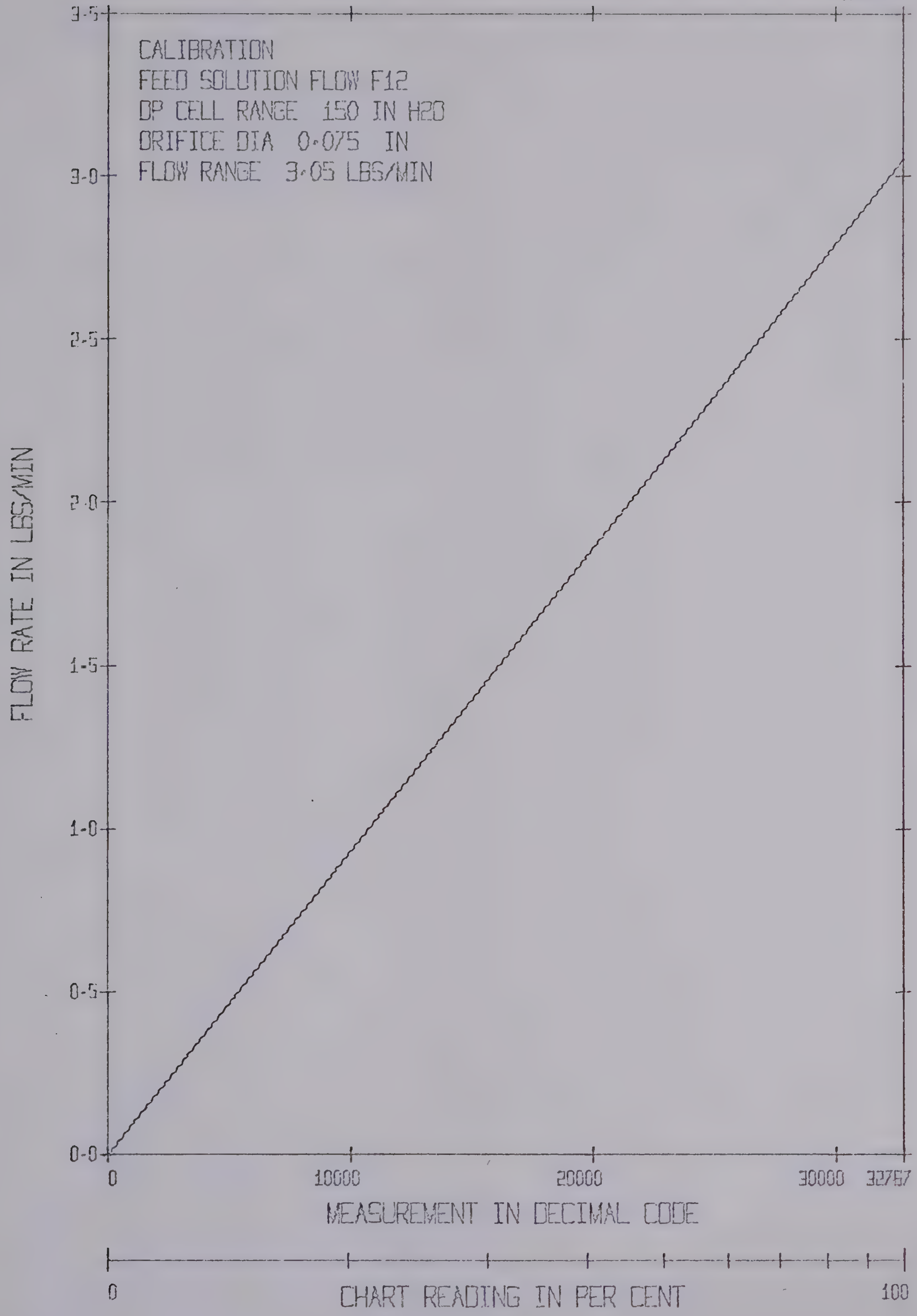

CALIBRATION
FEED SOLUTION FLOW F12
DP CELL RANGE 150 IN H2O
ORIFICE DIA 0.075 IN
FLOW RANGE 3.05 LBS/MIN
FLOW RATE IN LBS/MIN
0.0
0.5
1.0
1.5
2.0
2.5
3.0
3.5
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-10

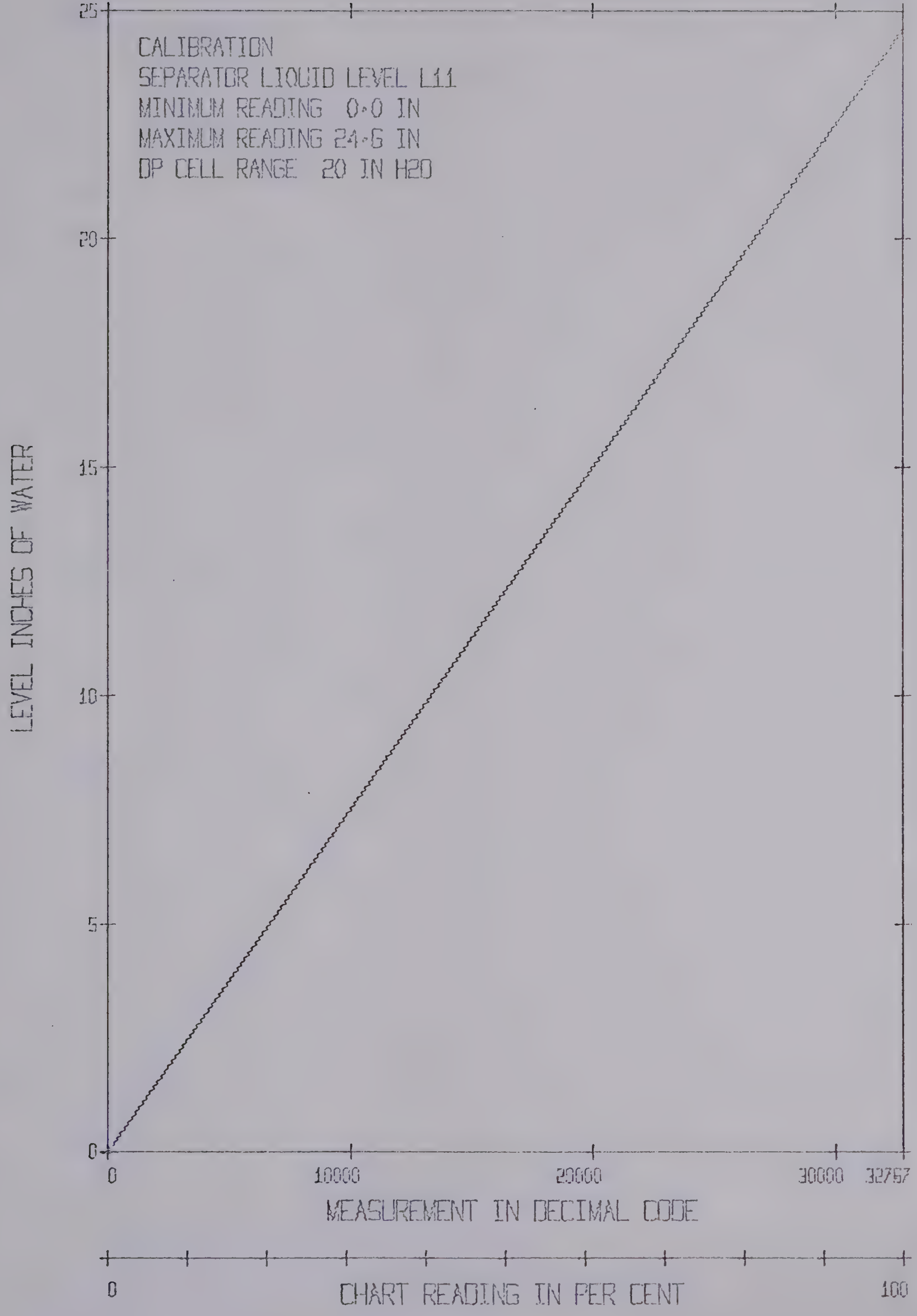

CALIBRATION
SEPARATOR LIQUID LEVEL L11
MINIMUM READING 0.0 IN
MAXIMUM READING 24.6 IN
DP CELL RANGE 20 IN H2O
LEVEL INCHES OF WATER
0
5
10
15
20
25
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-11

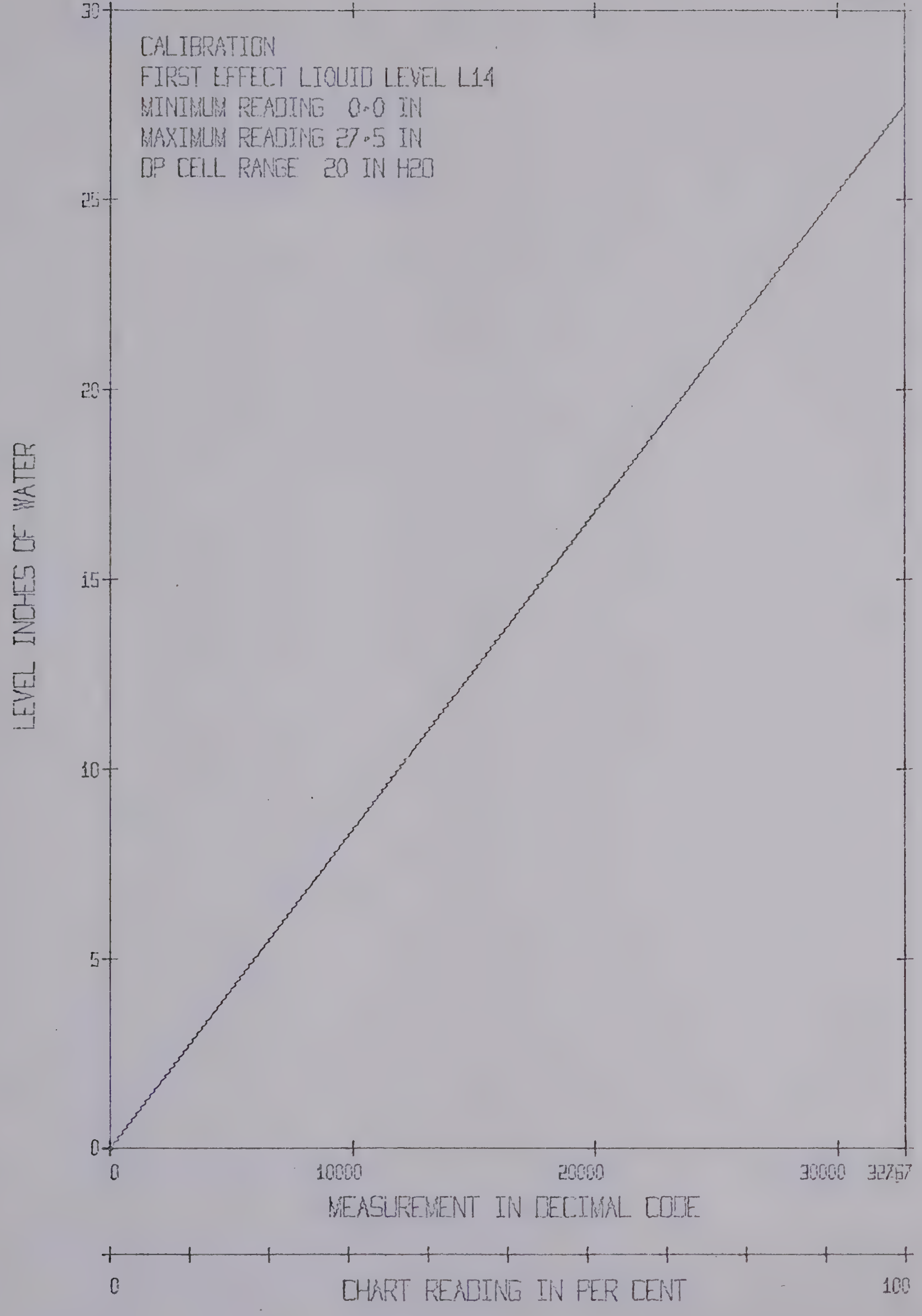

CALIBRATION
FIRST EFFECT LIQUID LEVEL L14
MINIMUM READING 0.0 IN
MAXIMUM READING 27.5 IN
DP CELL RANGE 20 IN H2O
LEVEL INCHES OF WATER
30
25
20
15
10
5
0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-12

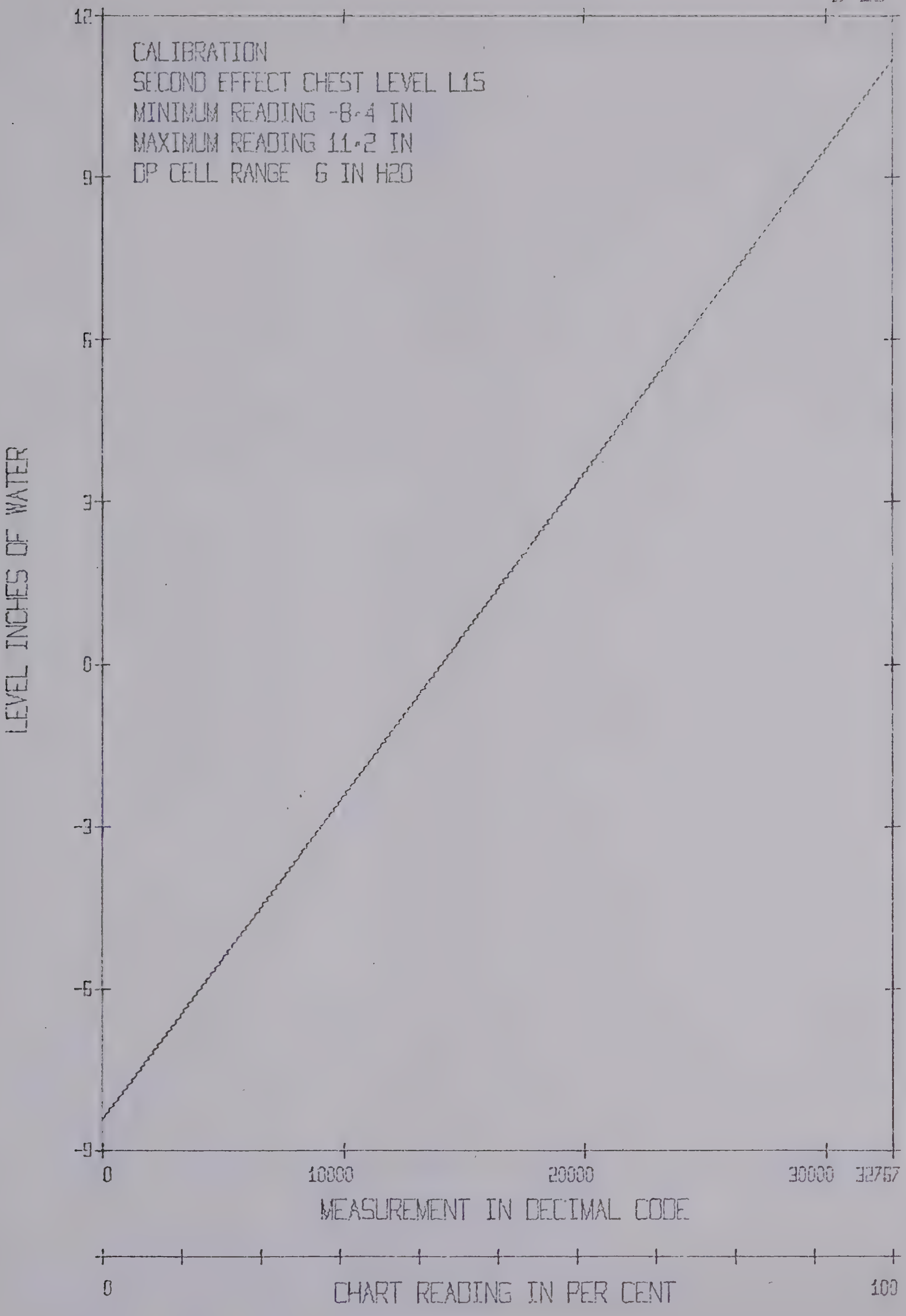

CALIBRATION
SECOND EFFECT CHEST LEVEL L15
MINIMUM READING -8.4 IN
MAXIMUM READING 11.2 IN
DP CELL RANGE 6 IN H2O
LEVEL INCHES OF WATER
12
9
6
3
0
-3
-6
-9
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-13

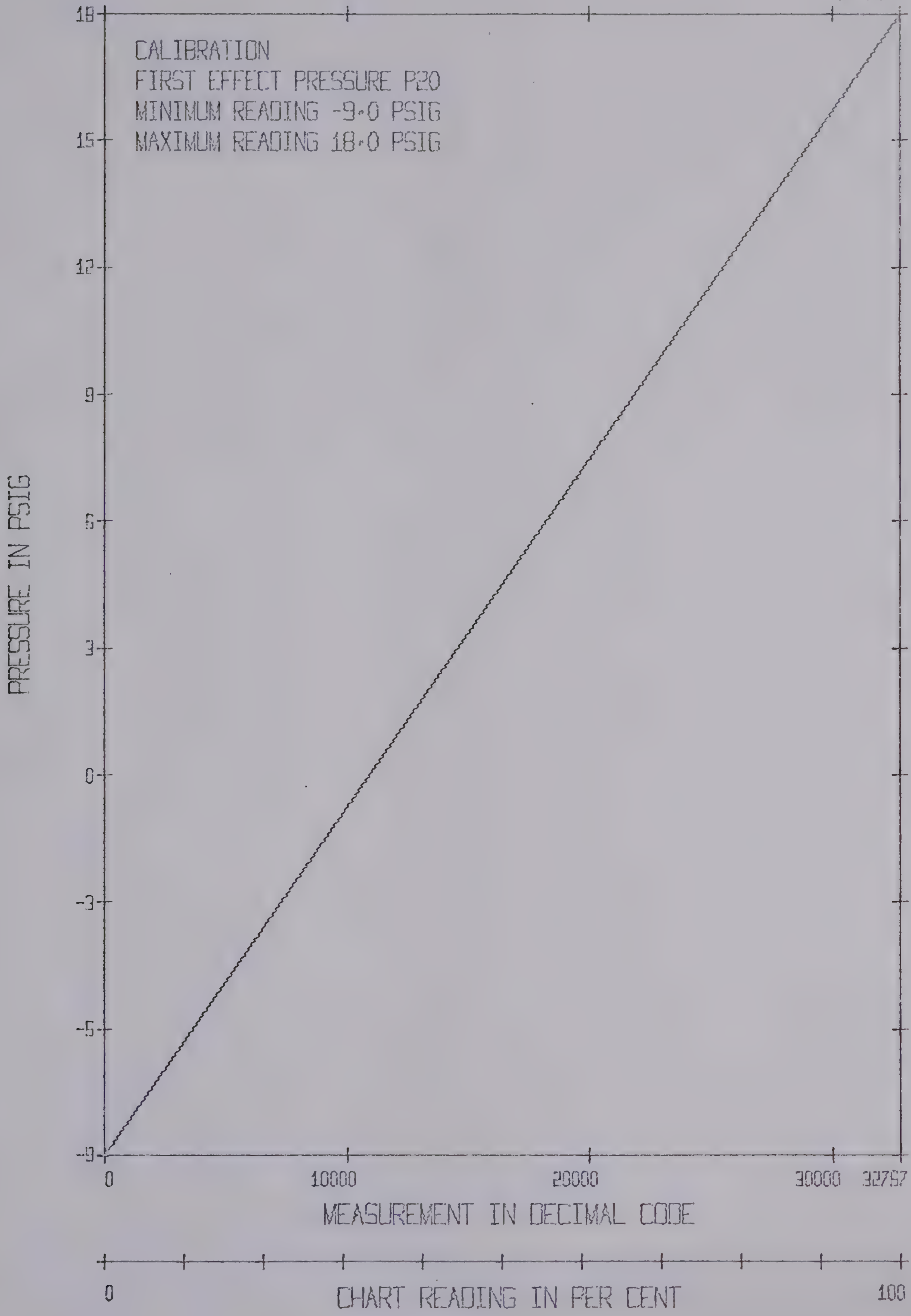

CALIBRATION
FIRST EFFECT PRESSURE P20
MINIMUM READING -9.0 PSIG
MAXIMUM READING 18.0 PSIG
PRESSURE IN PSIG
18
15
12
9
6
3
0
-3
-6
-9
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-14

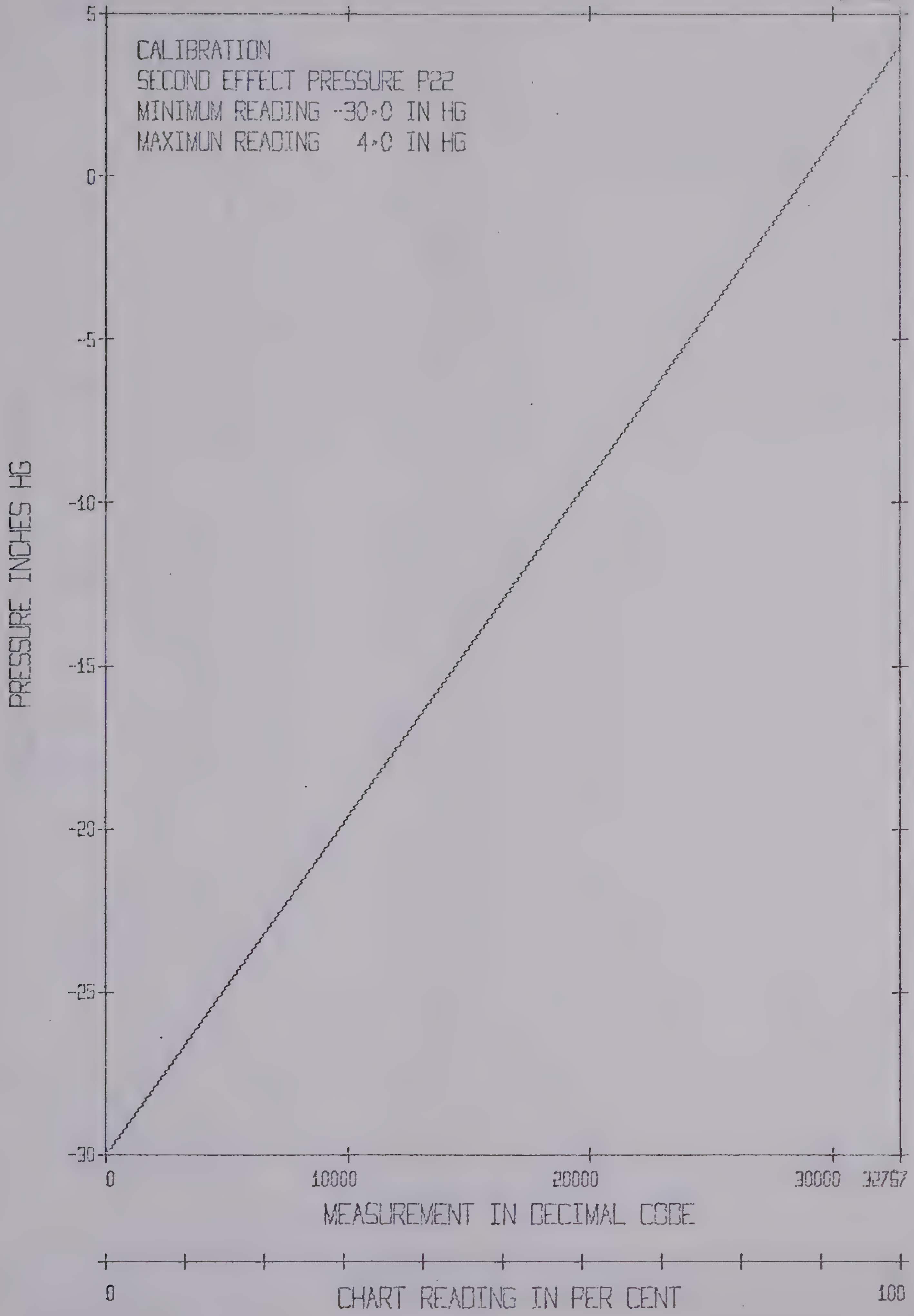

CALIBRATION
SECOND EFFECT PRESSURE P22
MINIMUM READING -30.0 IN HG
MAXIMUN READING 4.0 IN HG
PRESSURE INCHES HG
5
0
-5
-10
-15
-20
-25
-30
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-15

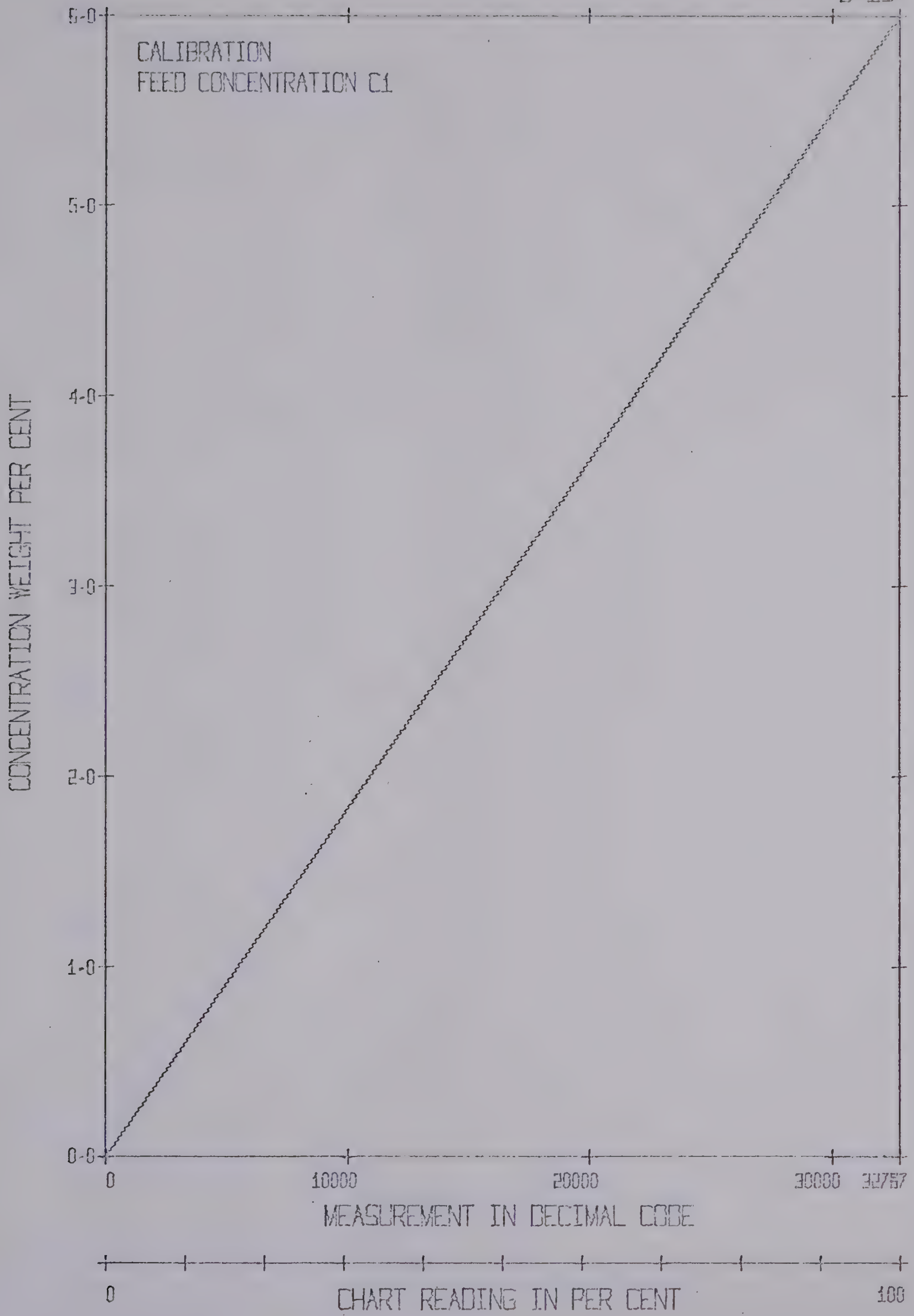

CALIBRATION
FEED CONCENTRATION C1
CONCENTRATION WEIGHT PER CENT
6-0
5-0
4-0
3-0
2-0
1-0
0-0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-16

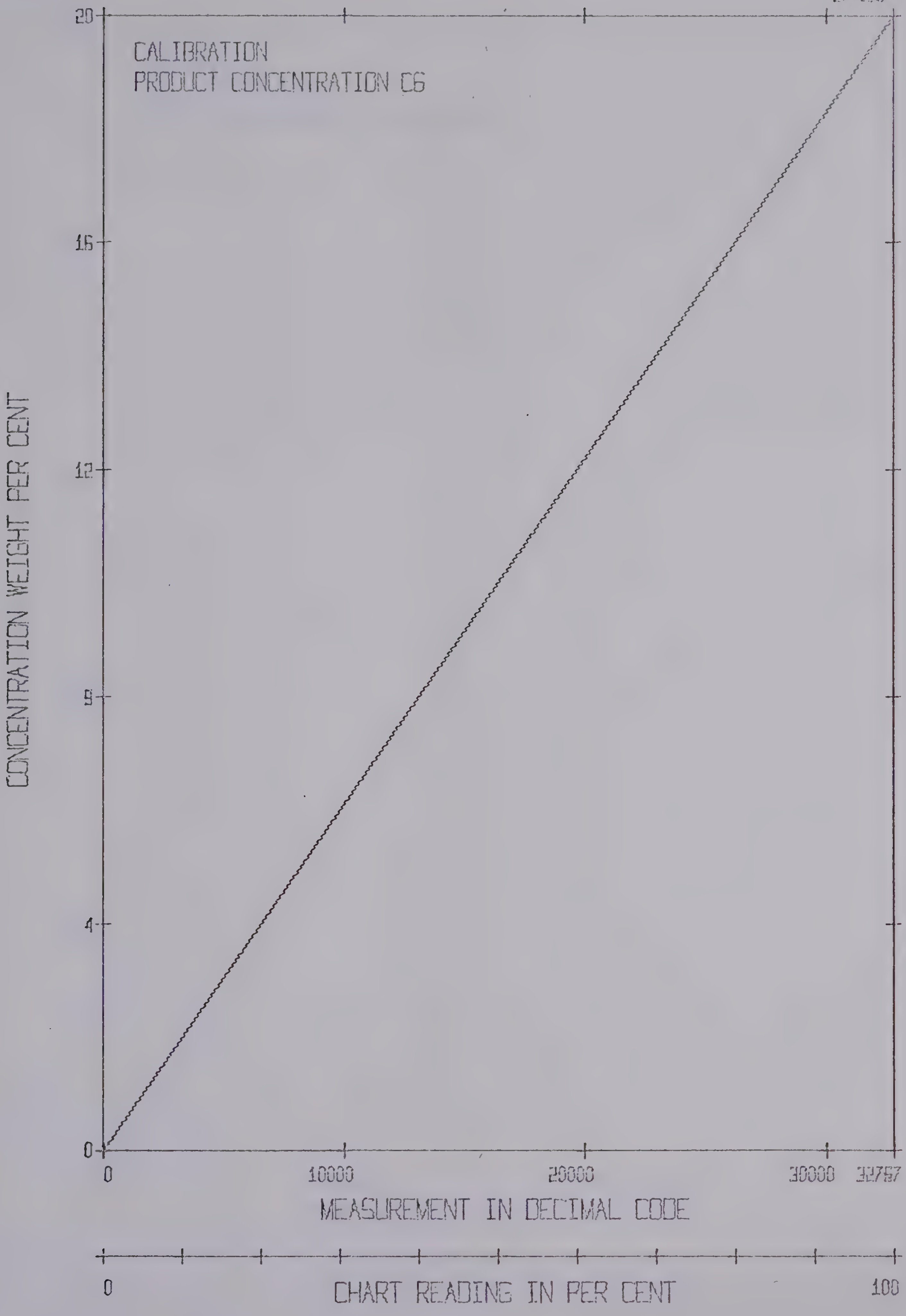
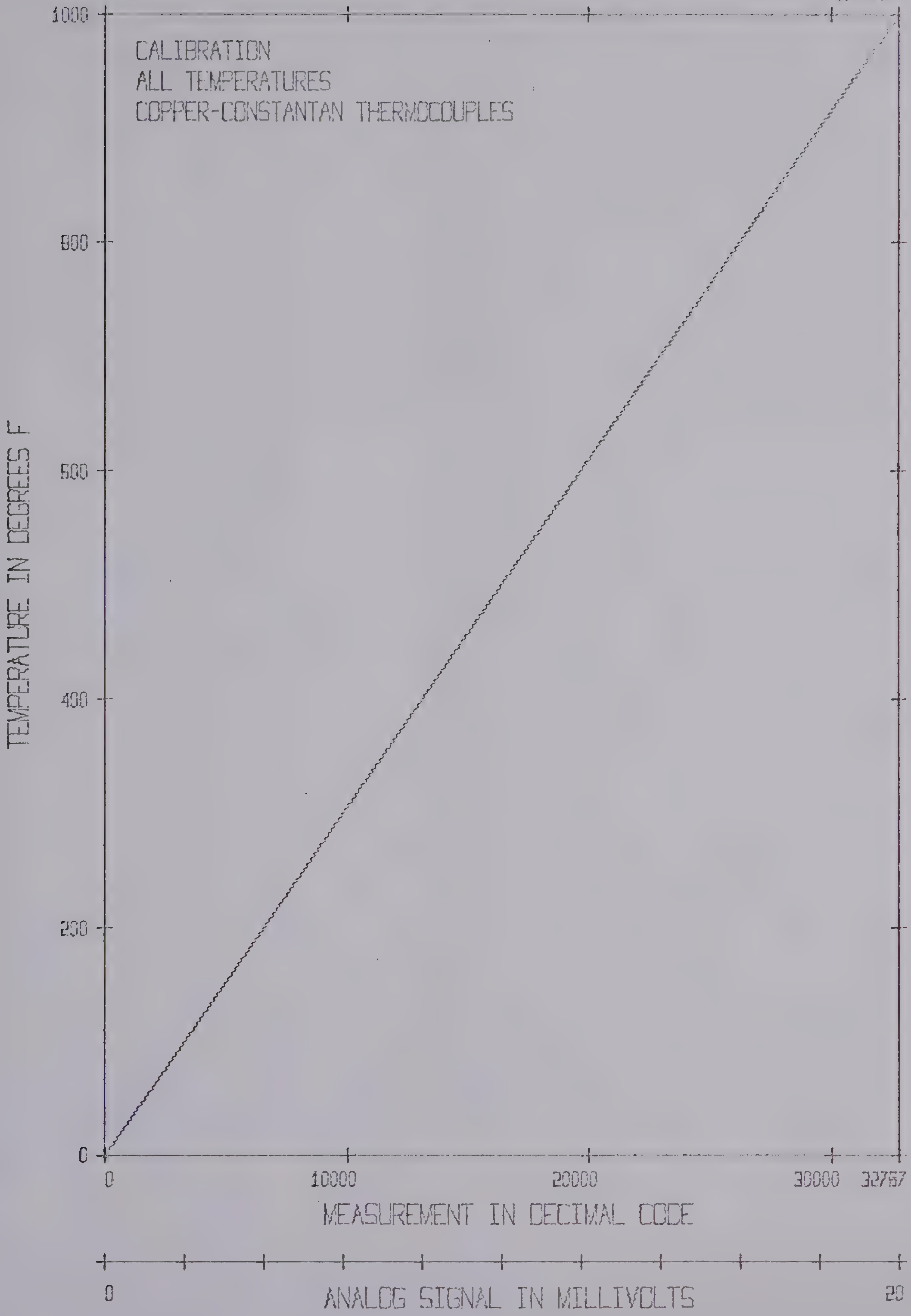

CALIBRATION
PRODUCT CONCENTRATION C6
CONCENTRATION WEIGHT PER CENT
20
16
12
8
4
0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
CHART READING IN PER CENT
100

FIGURE B-17

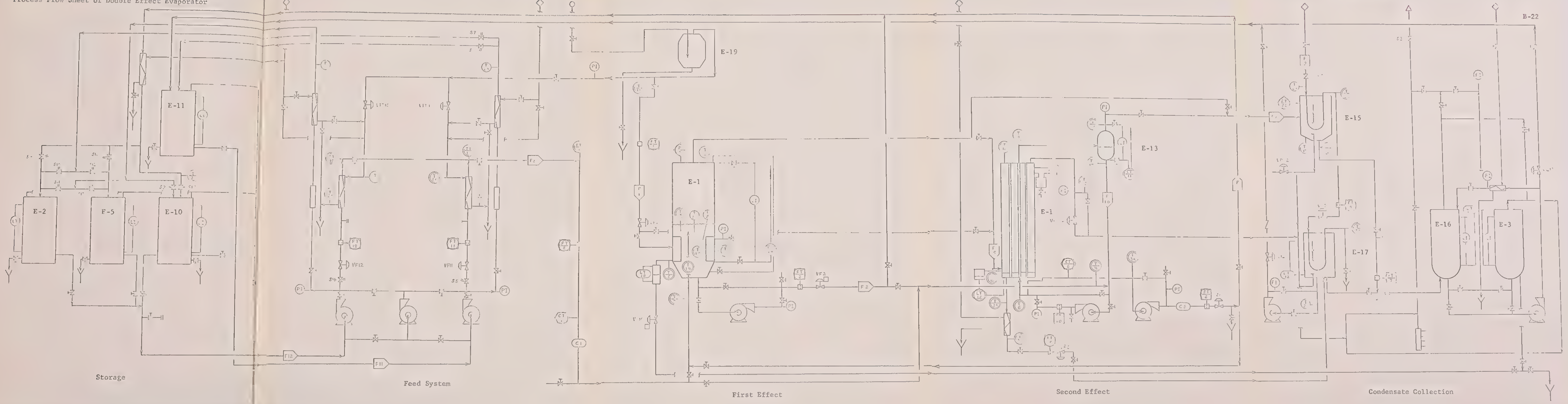


CALIBRATION
ALL TEMPERATURES
COPPER-CONSTANTAN THERMOCOUPLES
TEMPERATURE IN DEGREES F
1000
800
600
400
200
0
0
10000
20000
30000
32767
MEASUREMENT IN DECIMAL CODE
0
ANALOG SIGNAL IN MILLIVOLTS
20

FIGURE B-18

Figure B-19

Process Flow Sheet of Double Effect Evaporator

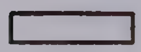


Legend to Figure B-19

Note: The DDC facilities allow for a wide range of control schemes to be implemented readily. The controller configuration employed in this study is shown on Figures 3.1 and 3.2.

Vessels:	E-1	first effect main body
	E-2	storage tank
	E-3	service side condensate tank
	E-5	storage tank
	E-10	storage tank
	E-11	water tank
	E-13	separator
	E-14	second effect main body
	E-15	condenser
	E-16	process side condensate tank
	E-17	rundown tank
	E-19	damper

 heat exchanger

 cartridge filter

 vacuum manifold

Table B-3


stream identification


hand valve


solenoid valve


check valve


pneumatic control valve


pneumatic control valve with valve positioner


orifice


cold water header

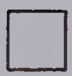

steam header


vacuum header

pressure gauge

sight glass

thermocouple


flow transmitter


level transmitter

Table B-3 continued

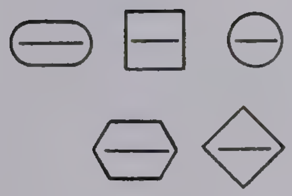
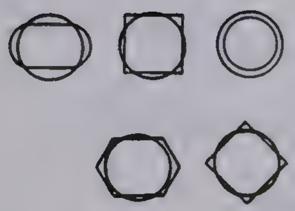
	pressure transmitter
	concentration transmitter
	a horizontal bar indicates the existence of a panel mounted recorder for the variable concerned
	an outer circle indicates that the variable can be read by the computer
Valve Labels	solenoid valves are identified as per reference (32)
	pneumatic control valves are identified by the letter V followed by the variable regulated as per Table B-2

Table B-3 continued

Appendix C

Control and Filter Algorithms

Digital Controllers

proportional	V = KP*E + B
integral	V = KP*KI*E + REST + B
proportional-integral	V = KP*E + KP*KI*E + REST + B

Digital Filters

exponential filter

$$PVO = (1-FF)*PVI + FF*PVOP$$

$$\ln(FF)*TI = -DT \; \dagger$$

Union filter

$$PVO = PVI \quad \text{when } |PVI-PVOP| \leq D$$

$$PVO = PVOP + (\text{sign of } PVI)*D \quad \text{when } |PVI-PVOP| > D$$

$$D = M \quad \text{when } |SP-PVOP| \leq M$$

$$D = |SP-PVOP| \quad \text{when } |SP-PVOP| > M$$

† recommended for use $0.8 \leq FF \leq 1.0$ (see page D-11)

Table C-1

Definition of Symbols

B	output bias
DT	poll time interval
E	effective error = filtered input - setpoint + error bias
FF	filter factor, positive constant, $0 < FF \leq 1$
KI	DT/TI
KP	proportional constant
M	quasi filter factor, constant defining allowable distance $\lvert PVO - PVOP \rvert$
PVI	raw value of process variable
PVO	filtered value of process variable
PVOP	filtered value of process variable at previous poll
REST	sum of previous integral contributions to V
SP	setpoint
TI	integral (reset) time as used in continuous controllers
V	control loop output

Table C-1 continued

Calculations

Symbols as defined on Figure 5.1

Feedforward Controller

Derivation of Equation (5-9)

Given: process model in form of equations (5-7) and (5-8)

Required: relation WST/CI

Solution:

control criterion CO = constant

$$\frac{dCO}{dt} = 0 \tag{C-1}$$

consider changes in CI independently, assume

$$WI = \text{constant} \tag{C-2}$$

substitute (C-1) and (C-2) into (5-8)

$$0 = \overline{WI}*C2 - 0.5*\overline{STEC}*WST*C2 - \overline{WI}*\overline{CO} + STEC*WST*\overline{CO} \tag{C-3}$$

linearize (C-3) by using perturbation techniques

$$C2' = \frac{0.5*STEC*\overline{C2} - STEC*\overline{CO}}{\overline{WI} - 0.5*STEC*\overline{WST}} * WST' \tag{C-4}$$

substitute (C-2) into (5-7)

$$\frac{dC2}{dt} = \frac{1}{H1}*\overline{WI}*CI - \frac{1}{H1}*\overline{WI}*C2 + \frac{STEC}{2*H1}*WST*C2 \tag{C-5}$$

linearize (C-5) by using perturbation techniques

$$H1*\frac{dC2'}{dt} = (\overline{WI})*CI' + (o.5*STEC*\overline{WST} - \overline{WI})*C2' + (o.5*STEC*\overline{C2})*WST' \qquad (C\text{-}6)$$

substitute (C-4) into (C-6) to obtain (5-9)

Derivation of Equation (5-10)

Given: equation (5-9) and steady state values from Table A-1

Required: transfer function WST(s)/CI(s)

Solution:

STEC = (O1 + O2)/WST = 1.67

C2 = WI*CI/B1 = 4.85%

substitute steady state values in equation (5-9)

$$\frac{dWST'}{dt} = -0.060*CI' - 0.189*WST' \qquad (C\text{-}7)$$

apply Laplace transformation to obtain (5-10)

Derivation of Equation (5-11)

Given: equation (5-10) and steady state values from Table A-1

Required: transfer function WST(s)/WI(s)

Solution:

assume first order lag system with time constant equal to that of equation (5-10)

to determine gain factor, assume constant steam economy which is realistic according to table 7.2

STEC = (O1 + O2)/WST = 1.67

fraction of feed evaporated at initial state

(O1 + O2)/WI = 0.67

fraction of feed evaporated at disturbed state

delta(O1 + O2)/delta(WI) = 0.67

= STEC*delta(WST)/delta(WI)

it follows that

delta(WST) = 0.67*delta(WI)/STEC = 0.40*delta(WI)

where delta designates the deviation from steady state

Derivation of Equation (5-12)

Given: steady state values from table A-1

Required: transfer function WST(s)/B1(s)

Solution:

analogous to the previous derivation
assume constant steam economy

ratio of quantity evaporated to first effect bottoms flow at initial state

(O1 + O2)/B1 = 1.01

at disturbed state

delta(O1 + O2)/delta(B1) = 1.01

= STEC*delta(WST)/delta(B1)

it follows that

delta(WST) = 1.01*delta(B1)/STEC = 0.61*delta(B1)

Inferential Controller

Derivation of Equation (5-13)

Given: equation (5-7)

Required: equation (5-13)

Solution:

linearize (5-7) by using perturbation techniques

$$\frac{dC2'}{dt} = CST1C*CI' + CST9W*WI' + CST2A*C2' + CST2W*WST' \qquad (C\text{-}8)$$

where

$CST1C = \overline{WI}/H1$

$CST2A = CST1B = CST9C = (STEC*\overline{WST} - 2*\overline{WI})/(2*H1)$

$CST9W = (\overline{CI} - \overline{C2})/H1$

$CST2W = STEC*\overline{C2}/(2*H1)$

apply Laplace transformation to (C-8) to obtain (5-13)

Derivation of Equation (5-14)

Given: equation (5-8)

Required: equation (5-14)

Solution:

linearize (5-8) by using perturbation techniques

$$\frac{dCO'}{dt} = CST4C*C2' + CSTTW*WI' + CST5D*CO' + CST5W*WST' \qquad (C\text{-}9)$$

where

$$CST4C = (2*\overline{WI} - STEC*\overline{WST})/(2*H2)$$

$$CSTTW = (\overline{C2} - \overline{CO})/H2$$

$$CST5D = CST4E = CSTTF = (STEC*\overline{WST} - \overline{WI})/H2$$

$$CST5W = (2*STEC*\overline{CO} - STEC*\overline{C2})/(2*H2)$$

apply Laplace transformation to (C-9) to obtain (5-14)

Building Blocks on Figure 5.4

The building blocks on Figure 5.4 represent the following expressions

block 1	CST1C/(s - CST1B)
block 2	CST2W/(s - CST2A)
block 9	CST9W/(s - CST9C)
block 4	CST4C/(s - CST4E)
block 5	CST5W/(s - CST5D)
block 10	CSTTW/(s - CSTTF)

Analytical Solution of Individual Equations

In the solution of equations (5-13) and (5-14), each term was treated separately and the results added in accordance with the block diagram shown in Figure 5.4.

Example: block 1

The equivalent differential equation of the first term in equation (5-13) is

$$\frac{dB}{dt} = CST1B*B + CST1C*CI \tag{C-10}$$

The solution was obtained via Laplace transformation, considering initial conditions, B (o).

$$s*B(s) - B(o) = CST1B*B(s) + CST1C*\frac{CI}{s} \tag{C-11}$$

$$B(s) = \frac{B(o)}{s - CST1B} + \frac{CST1C*CI}{s(s - CST1B)} \tag{C-12}$$

$$B(t) = B(o)*\exp(CST1B*t) + \frac{CST1C*CI}{-CST1B} * (1 - \exp(CST1B*t)) \tag{C-13}$$

The value of the constants was calculated using steady state readings at the beginning of each run. The value of the initial conditions B(o) was then determined from the steady state value of CI obtained from the process variable table and the relation required by equation (C-10) as follows

$$B(o) \equiv - \frac{CST1C*CI}{CST1B} \tag{C-14}$$

The value of t was constant and equal to the computation interval H. The value of CI was obtained directly from the process variable table every time a computation was called for. The values of B(t) at each sampling interval during the transient were evaluated from a difference equation based on (C-13) of the following general form

$$B\left[(n+1)H\right] = B\left[nH\right] * K1 + CI\left[nH\right] * K2 \tag{C-15}$$

where

$$K1 = \exp(CST1B*H)$$

$$K2 = - \frac{CST1C}{CST1B} * (1 - \exp(CST1B*H))$$

Calculations were performed by subroutine BLUE of program COLOR.

Appendix D

COMPLETE LIST OF VARIABLE RECORDS

COS - CURRENT OUTPUT STATION
MPX - RELAY MULTIPLEXER POINT

LOOP	NAME	COS	MPX	VARIABLE DESCRIPTION
0125	L15		28	SECOND EFFECT CHEST LEVEL
0126	F7		42	SECOND EFFECT OVERHEAD FLOW
0127				DATA ACCUM. FROM LOOP 0152
0128				DATA ACCUM. FROM LOOP 0151
0129			35	ARTIFICIAL READING OF PRODUCT CONC.
0130			32	ARTIFICIAL READING OF FEED CONC.
0131				DATA ACCUM. FROM LOOP 0132
0132	F1	III-6	0	STEAM FLOW TO FIRST EFFECT
0133	F9	II-6	16	COOLING WATER FLOW TO CONDENSER
0134	P22	III-3	25	SECOND EFFECT PRESSURE
0135	P20	III-4	41	FIRST EFFECT PRESSURE
0136	L14		22	FIRST EFFECT LIQUID LEVEL
0137	L11		19	SEPARATOR LIQUID LEVEL
0138	TT10	II-5	84	TEG FEED SOLUTION TEMPERATURE
0139	F12	II-4	9	TEG FEED SOLUTION FLOW
0140				RATIO OF FEED FLOWS
0141	F11	II-3	6	FEED WATER FLOW
0142	TT11	II-2	83	FEED WATER TEMPERATURE
0143	C1		32	FEED CONCENTRATION
0144	C6		35	PRODUCT CONCENTRATION
0145			32	FEEDFORWARD COMPENSATION FOR CONC.
0146				DATA ACCUM. FROM LOOP 0137
0147				DATA ACCUM. FROM LOOP 0144
0148				DATA ACCUM. FROM LOOP 0136
0149	F8		12	TOTAL FEED FLOW
0150	F5		38	FIRST EFFECT OVERHEAD FLOW
0151	F2	III-2	39	FIRST EFFECT BOTTOMS FLOW
0152	F6	III-1	40	PRODUCT FLOW FROM SECOND EFFECT
0153	RBT2		65	BRIDGE UNBALANCE MEASUREMENT
0154	RBT1		64	REFERENCE VOLTAGE
0155	T1		100	COOLING WATER AT CONDENSER OUTLET
0156	T2		90	FIRST EFFECT VAPOUR SPACE
0157	T4		92	SOLUTION TO SECOND EFFECT
0158	T5		88	STEAM CONDENSATE FROM FIRST EFFECT
0159	T7		91	FEED TO FIRST EFFECT
0160	T10		97	STEAM TO SECOND EFFECT
0161	T11		103	CONDENSER CONDENSATE
0162	T12		96	SEPARATOR VAPOUR
0163	T15		86	STEAM SUPPLY TO EVAPORATOR
0164	T19		89	LIQUID IN FIRST EFFECT
0165	T28		98	SECOND EFFECT STEAM CONDENSATE
0166	T30		101	RUNDOWN TANK COOLING WATER OUT
0167	T34		79	PRODUCT FROM SECOND EFFECT
0168			12	FEEDFORWARD COMPENSATION FOR FLOW
0169				DUMMY LOOP FOR ADDITION OF SIGNALS

TABLE D-1

EXAMPLE OF PROCESS VARIABLE TABLE

COLUMN 1 - VARIABLE RECORD IDENTIFICATION
COLUMNS 2 TO 10 - VARIABLE RECORD CONTENTS

```
012501  E21E  374D+00028  3C00  9540+03920-00840+15944+15731
012502+32767+32767  3220+32767-32768  2205+32767-32768  3220
012503  0150+19648  C0A1  0000+18720-17920  0001  0200  0000
012504  0000+00000
012601  E00D  4F5D+00042  3200  8440+07500+00000+14754+00000
012602+32767+32767  3220
012701  8625  6D08  0152  C024  03B2  0202  4ACF  48AF  49CF
012702  4AA3  49B3  49A0  49D4  47EA  4820  480B  4802  4890
012703  48A7  49D5  49F1  4A4A  49D3  4960  494F  49A6  4A30
012704  4935  491D  4A31  4957  4970  49BE  49C4  49C7  4B30
012801  8625  6E08  0151  C000  03B2  0203  4749  47B2  4747
012802  4732  45E3  45FC  4662  470A  46A6  46E1  46FA  4749
012803  46E4  4711  46B9  45AE  45B3  4570  45E5  466E  4674
012804  46C4  4778  474D  46D2  46B2  467D  4613  4643  467C
012901  D21E  374D+00035  0200  9141+04000+00000+16963+00000
012902+32767+32767  3220+32767-32768  3220+32767-32768  3220
012903  0000+16850  1000  0000+00000+00000  0000  0080  0000
012904  0000+00000
013001  D21E  374D+00032  0200  9141+01200+00000-00009+00000
013002+32767+32767  3220+32767-32768  3220+32767-32768  3220
013003  0000+16384  1000  0000+00000+00000  0000  0080  0000
013004  0000+00000
013101  8625  6808  0132  C00B  03B1  0239  52CC  531B  52D5
013102  505A  52CE  52C6  52CA  533C  52C2  52B9  52B5  52F6
013103  53D5  52E5  525C  526F  530B  5326  5353  531B  52DB
013104  52F7  5317  5313  5360  528E  5290  52A1  5296  52BD
013201  C21E  245D+00000  0000  84C1+07160+00000+20398+20142
013202+32767+32767  3220+32767-32768  3220+32767-32768  3220
013203  1D78+24200  C3C0  0000+24328+16384  0200  0040  0000
013204  0000+00000
013301  E21E  235D+00016  3400  94E1+10080+00000+12279+32507
013302+32767+32767  3220+32767-32768  3220+32767-32768  F220
013303  1578+32767  C0C0  0000+32767-32768  0200  0040  0000
013304  0000+00000
013401  E21E  114D+00025  3000  96C0+06800+03000+18263+18311
013402+32767+32767  3220+32767+00000  3A20+32767+00000  3220
013403  1A78+26560  C0C0  0000+19264+14336  0001  4C00  0000
013404  0000+00000
013501  C41E  374D+00041  0000  96C1+10960-01830+20096+20210
013502+32767+32767  3220+32767+00000  3A20+32767+00000  3220
013503  1B78+27518  00C0  0000+24640-32768  0080  0E00  0000
013504  8010-00314
```

TABLE D-2

EXAMPLE OF PROCESS VARIABLE TABLE

```
013601  E21E  314D+00022  3C00  95C1+05500+00000+13411+13107
013602+32767+32767  3220+32767+00000  3A20+32767+00000  3220
013603  0151+18080  C4A1  0000+17815+10560  0002  0070  0000
013604  0000+00000
013701  E21E  354D+00019  3C00  95C1+04920+00000+09283+09151
013702+32767+32767  3220+32767+00000  3A20+32767+00000  3220
013703  0152+18944  C0A1  0000+18549+23168  0002  0180  0000
013704  0000+00000
013801  C21E  33CC+00084  0000  A2C1+20000+00000+07014+06389
013802+32767+32767  3220+32767+00000  3220+32767+00000  3220
013803  1478+29175  C0C0  0000+29800-32768  0100  0080  0000
013804  0000+00000
013901  D21E  005D+00009  0400  84C0+06100+00000+12899+11393
013902+32767+32767  3220+32767+00000  3A20+32767+00000  3220
013903  1378+13256  C3C0  0000+14277+10112  0007  0034  0000
013904  0000+00000
014001  8A1E  007C+00313  0000  84C0+06100+00000+15037+00000
014002+32767+32767  3220+32767+00000  3A20+32767+00000  3220
014003  0141+21700  80A1  0260+00000+00000  0000  2200  0000
014004  0000+00000
014101  D21E  005D+00006  0400  84C0+10200+00000+22746+20938
014102+32767+32767  3220+32767+00000  3A20+32767+00000  3220
014103  1278+21415  C3C0  0000+21671+27712  0003  006B  0000
014104  0000+00000
014201  C21E  32CC+00083  0000  A2C0+20000+00000+06405+06389
014202+32767+32767  3220+32767+00000  3220+32767+00000  3220
014203  1178+25576  C0C0  0000+25592-32768  0200  0080  0000
014204  0000+00000
014301  821E  366C+00304  0000  9100+01200+00000+16384+21845
014302+32767+32767  3220+32767-32768  3220+32767-32768  3220
014303  0140+08704  00A3  0000+08436-01536  0004  0080  0000
014304  0000+0000
014401  821E  376C+00297  0000  9140+04000+00000+16850+16842
014402+32767+32767  3220+32767-32768  3220+32767-32768  3220
014403  0132+19948  C0E1  FAB1+21251+06400  000C  0028  0000
014404  0000+00000
014501  A21E  336C+00304  3E80  9100+01200+00000+16342+16384
014502+32767+32767  3220+32767-32768  3220+32767-32768  3220
014503  0169+00022  D0E1  0000+00000+00000  0000  0044  0000
014504  0000+00000
014601  8625  6E08  0137  C014  03B3  0207  2464  2420  240A
014602  2427  2407  23DE  23FA  240F  2446  2452  2450  2448
014603  242E  2475  244B  2466  246D  2451  2430  2401  23F9
014604  23E0  23D1  23C1  2417  2435  2450  2464  2486  2468
014701  8625  6108  0129  C015  03B2  0236  4216  4250  4214
014702  4214  4228  4225  4223  4216  4200  41F8  41F3  4216
```

TABLE D-2 CONTINUED

EXAMPLE OF PROCESS VARIABLE TABLE

```
014703  422D  4237  3FE4  4007  4045  4018  4009  4045  402C
014704  402F  402C  400E  4004  401B  4018  403E  3FFF  420F
014801  8625  6F08  0136  C013  03B3  0208  3461  347E  3524
014802  35BB  3591  350A  343C  3390  3333  342D  3464  3460
014803  3634  35CD  348B  334E  342B  3519  356E  34F6  3509
014804  3536  359A  3544  3494  33F8  32FB  325B  331F  33FC
014901  D21E  105D+00012  0600  8441+17440+00000+17465+16910
014902+32767+32767  3220+32767-32768  3220+32767-32768  3220
014903  0139+12548  C0E1  0000+17543+32192  0002  0240  0000
014904  0000+00000
015001  D21E  115D+00038  0800  8441+06460+00000+20028+21015
015002+32767+32767  3220+32767-32768  3220+32767-32768  3220
015003  1C78+09396  C3C0  0000+08415-20416  0001  0080  0000
015004  0000+00000
015101  D21E  005D+00039  0800  8441+12480+00000+17898+18025
015102+32767+32767  3220+32767-32768  3220+32767-32768  3220
015103  1978+27521  C380  0000+27575+22080  0002  0030  0000
015104  0000+00000
015201  D21E  105D+00040  0800  8440+05340+00000+18839+18841
015202+32767+32767  3220+32767-32768  3220+32767-32768  3220
015203  1878+11822  C380  0000+12471+06656  0003  0280  0000
015204  0000+00000
015301  E00D  6C0C+00065  3200  9700+04000+00000+09440+00000
015302+32767+32767  3220
015401  EF16  LB0C+00064  3200  9720+04000+00000+13186+00000
015402+32767+32767  3220+32767-32768  3220+32767-32768  3220
015403  0153+00000  F320
015501  E00D  40CC+00100  3A00  A200+20000+00000+03641+00000
015502+32767+32767  3220
015601  E00D  41CC+00090  3A00  A200+20000+00000+07597+00000
015602+32767+32767  3220
015701  E00D  42CC+00092  3A00  A200+20000+00000+06266+00000
015702+32767+32767  3220
015801  E00D  43CC+00088  3A00  A200+20000+00000+08126+00000
015802+32767+32767  3220
015901  E00D  44CC+00091  3A00  A200+20000+00000+06569+00000
015902+32767+32767  3220
016001  E00D  45CC+00097  3A00  A200+20000+00000+07585+00000
016002+32767+32767  3220
016101  E00D  46CC+00103  3A00  A200+20000+00000+04572+00000
016102+32767+32767  3220
016201  E00D  47CC+00096  3A00  A200+20000+00000+05692+00000
016202+32767+32767  3220
016301  E00D  48CC+00086  3A00  A200+20000+00000+10305+00000
016302+32767+32767  3220
016401  E00D  49CC+00089  3A00  A200+20000+00000+07615+00000
```

TABLE D-2 CONTINUED

EXAMPLE OF PROCESS VARIABLE TABLE

```
016402+32767+32767  3220
016501  E00D  4ACC+00098  3A00  A200+20000+00000+06922+00000
016502+32767+32767  3220
016601  E00D  4BCC+00101  3A00  A200+20000+00000+03720+00000
016602+32767+32767  3220
016701  E00D  4CCC+00079  3A00  A200+20000+00000+03335+00000
016702+32767+32767  3220
016801  E21E  325D+00012  3E80  8440+17440+00000+18833+20667
016802+32767+32767  3220+32767-32768  3220+32767-32768  3220
016803  0169+01776  D0E0  0000+00000+00000  0000  007C  0000
016804  0000+00000
016901  821E  2068+00360  0000  0F00+00000+00000+01850+00022
016902+32767+32767  3220+32767-32768  3220+32767-32768  3220
016903  0144-01828  D0E2  0000+00000+00000  0000  0080  0000
016904  0000+00000
```

TABLE D-2 CONTINUED

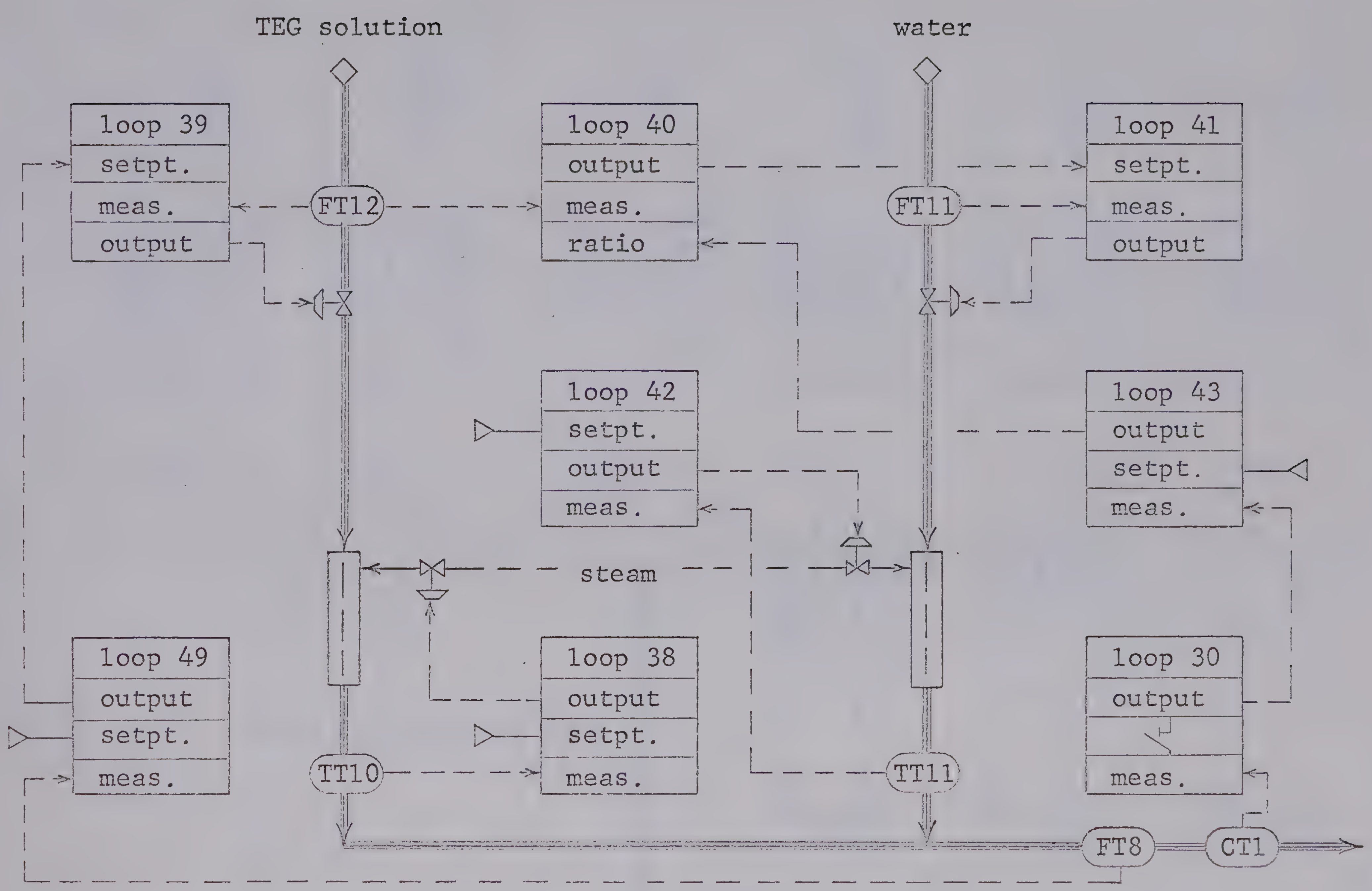


Figure D-1 Feed System Control Loops

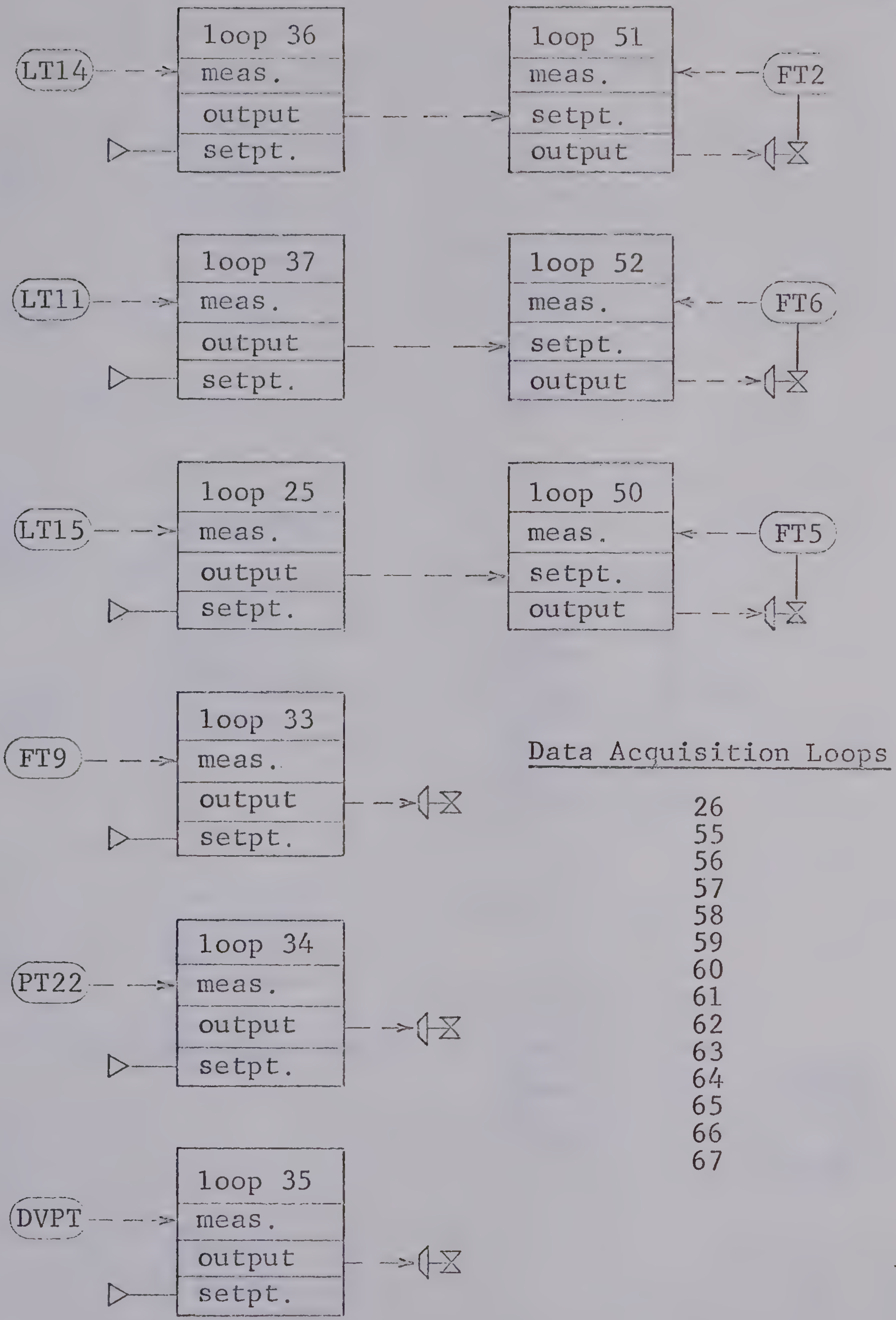
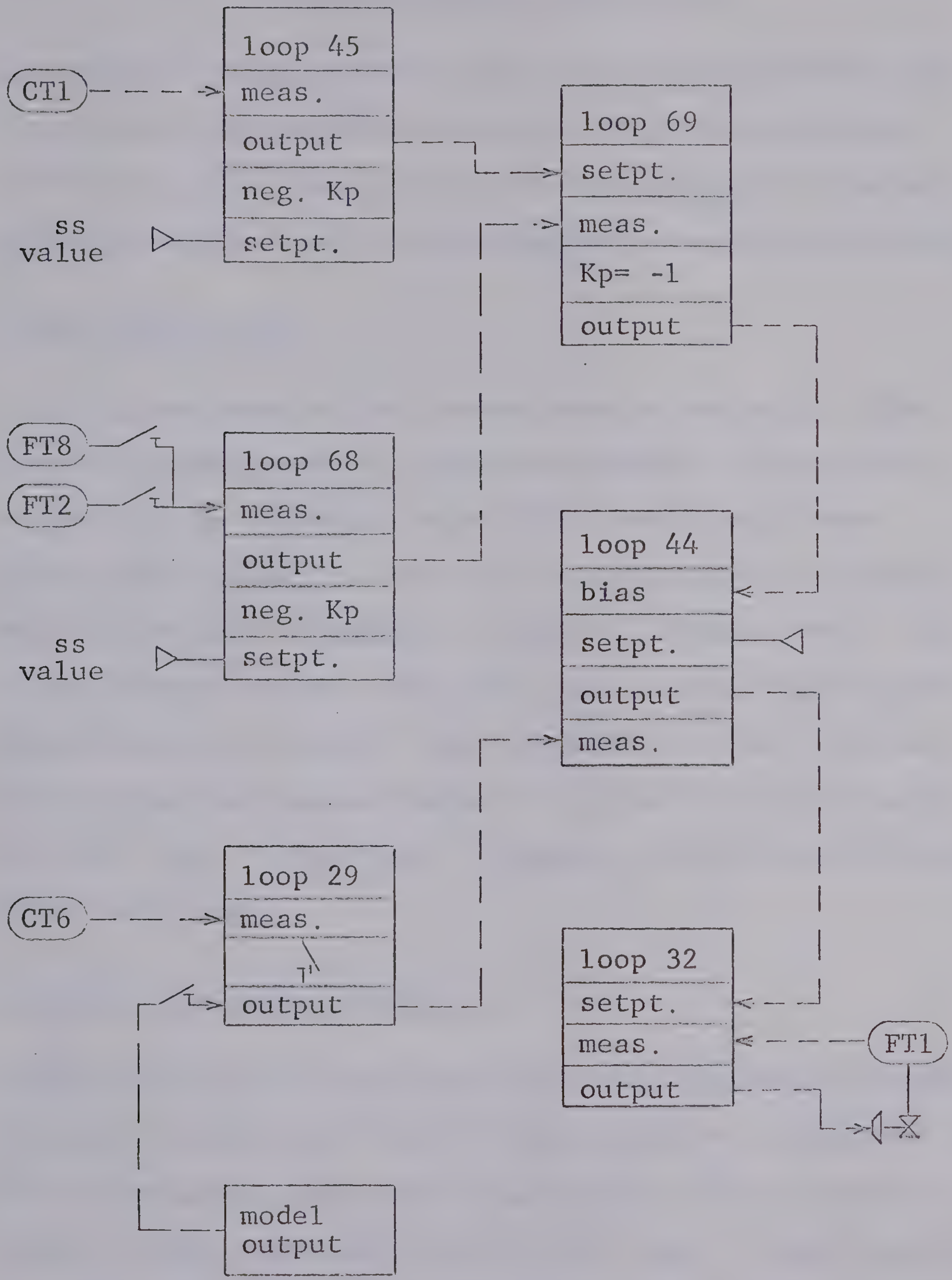


Figure D-2 Basic Control Loops



Product Concentration Control Loops

Figure D-3

Notes on Digital Controllers

All variable records were written in accordance with the Loop Record Builder Form available for this particular system (22,35). Their contents at one particular instant during the final day of experimentation are shown on Table D-2.

Feed Concentration

Loop 30 was inserted for convenience of operation. Whenever the refractometer functioned properly, loop 30 was placed in the automatic mode with a setpoint of zero and a gain of unity, such that the refractometer reading was transmitted unchanged to loop 43. In cases where the refractometer did not function, loop 30 was placed in the manual mode. The correct feed concentration was then entered from the process operator's console into the output of this loop. In this way, a reading was available to loop 43 at all times.

Differential Vapour Pressure

Difficulties were experienced with the differential vapour pressure transmitter. Thus it was necessary to operate loop 35 in the manual mode for a large part of the program. In order to make additional use of this loop, it was connected to the first effect pressure transmitter, such that this pressure reading was available from the console.

Feedback Control

Loop 44 was the basic feedback controller. Loop 29 was inserted for convenience of operation. It served two main purposes. Firstly, in case of instrument failure, it could receive the correct product concentration from the console, similar to loop 30. Secondly, when the process model was used for control purposes, process and model outputs could be recorded separately by loops 29 and 44, respectively. The output signal of the feedback concentration controller was transmitted to the setpoint of the steam flow controller.

Feedforward Control

The feedforward controller consisted of loops 45, 68, and 69. Steady state values of feed flow and concentration were supplied from the console as setpoints to loops 68 and 45, respectively. The measurements obtained from the process were filtered, and the steady state values subtracted from them, such that the perturbations, including dynamic compensation, were contained in the error signals. Dynamic compensation was achieved by the use of exponential filters, the filter constants being defined by the time constants in equations (5-10) and (5-11). Relation between filter factor FF, time constant TI , and poll time interval DT (35):

$$\ln(FF)*TI = -DT$$

In the present case TI = 5.3 minutes and DT = 8 seconds.
It follows that FF = 0.976.

Gain factors required by equations (5-10), (5-11), and (5-12)

were changed into range factors and entered into loops 45 and 68 as proportional constants as follows (see calibration curves in Appendix B for recorder ranges).

Feed concentration: gain required 0.314
range of steam flow recorder 3.58
range of concentr. recorder 6.0

$$\text{range factor} = \frac{\frac{0.314}{3.58} * 32767}{\frac{1.0}{6.0} * 32767} = 0.526$$

Feed flow: gain required 0.40
range of feed flow recorder 8.72

$$\text{range factor} = \frac{\frac{0.40}{3.58} * 32767}{\frac{1.0}{8.72} * 32767} = 0.97$$

First effect bottoms flow: gain required 0.61
range of recorder 6.24

$$\text{range factor} = \frac{\frac{0.61}{3.58} * 32767}{\frac{1.0}{6.24} * 32767} = 1.06$$

The two feedforward controller signals were added by loop 69 and transmitted to the steam flow controller via the output bias of loop 44. Thus the steam flow controller received its steady state signal from the steady state output of loop 44 and the compensation due to load disturbances from the bias of loop 44. In cases where feedforward control alone was used, the error of loop 44 was artificially zeroed out with the aid of loop 29.

LIST OF CONTROLLER AND FILTER CONSTANTS

VARIABLE DESCRIPTION	NAME	PROPOR-TIONAL CON-STANT	RESET TIME SEC.	FIL-TER TYPE	FILTER CON-STANT
STEAM FLOW TO FIRST EFFECT	F1	0.5	8		
FIRST EFFECT BOTTOMS FLOW	F2	0.375	500	UNION	256
FIRST EFFECT OVERHEAD FLOW	F5	1.0	2000	UNION	256
PRODUCT FLOW	F6	5.0	667	UNION	256
SECOND EFFECT OVERHEAD FLOW	F7			EXPON	0.78
TOTAL FEED FLOW	F8	4.5	1000	UNION	192
COOLING WATER FLOW TO CONDENSER	F9	0.5	8	EXPON	0.81
FEED WATER FLOW	F11	0.84	333	UNION	128
TEG FEED SOLUTION FLOW	F12	0.41	143	UNION	128
FEED CONCENTRATION	C1	1.0	2000		
PRODUCT CONCENTRATION	C6	0.31	684		
SECOND EFFECT PRESSURE	P22	304.0	2000	EXPON	0.75
SEPARATOR LIQUID LEVEL	L11	3.0	4000	EXPON	0.94
FIRST EFFECT LIQUID LEVEL	L14	0.875	4000	EXPON	0.94
TEG FEED SOLUTION TEMPERATURE	TT10	1.0	32		
FEED WATER TEMPERATURE	TT11	1.0	16		

TABLE D-3

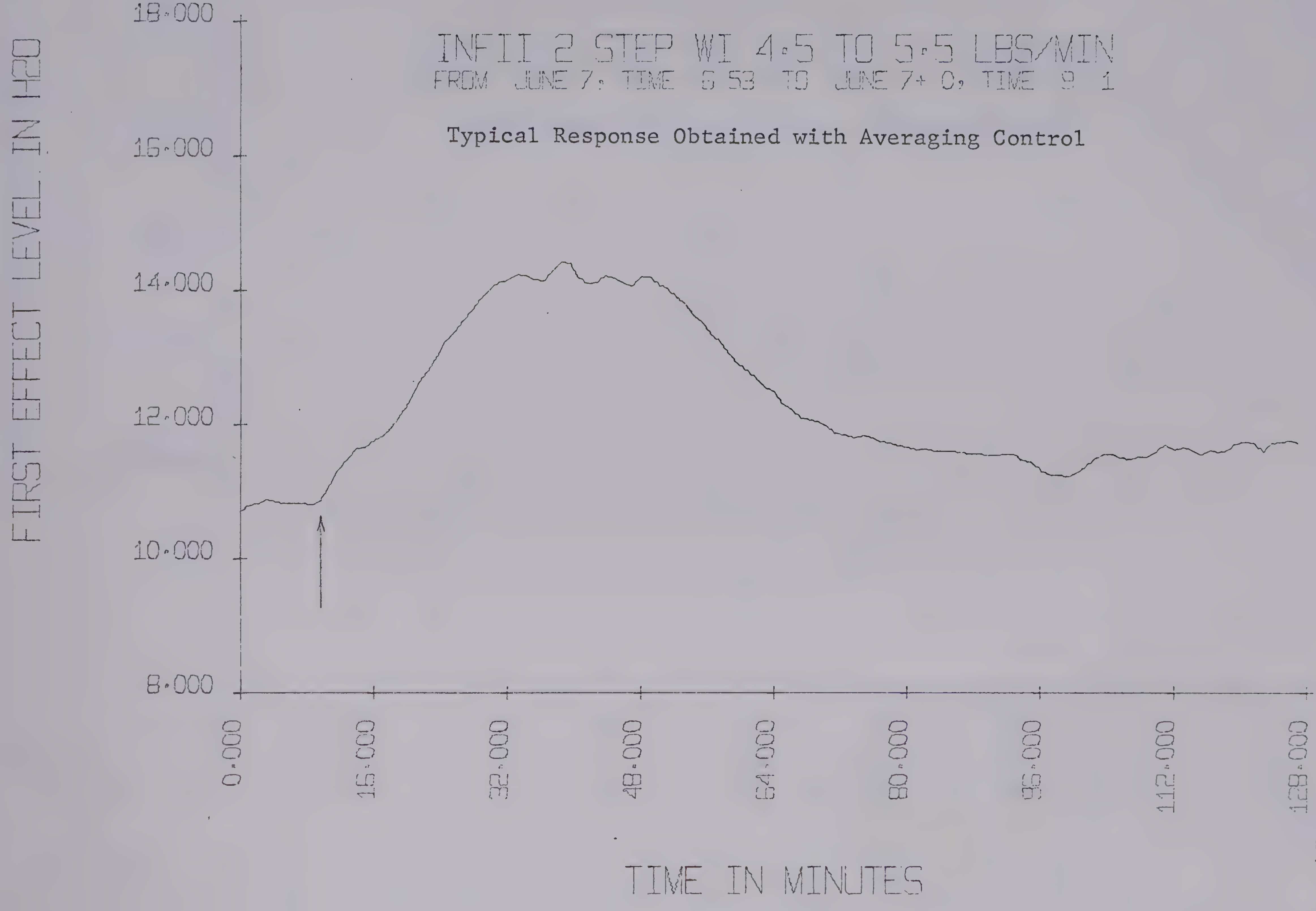
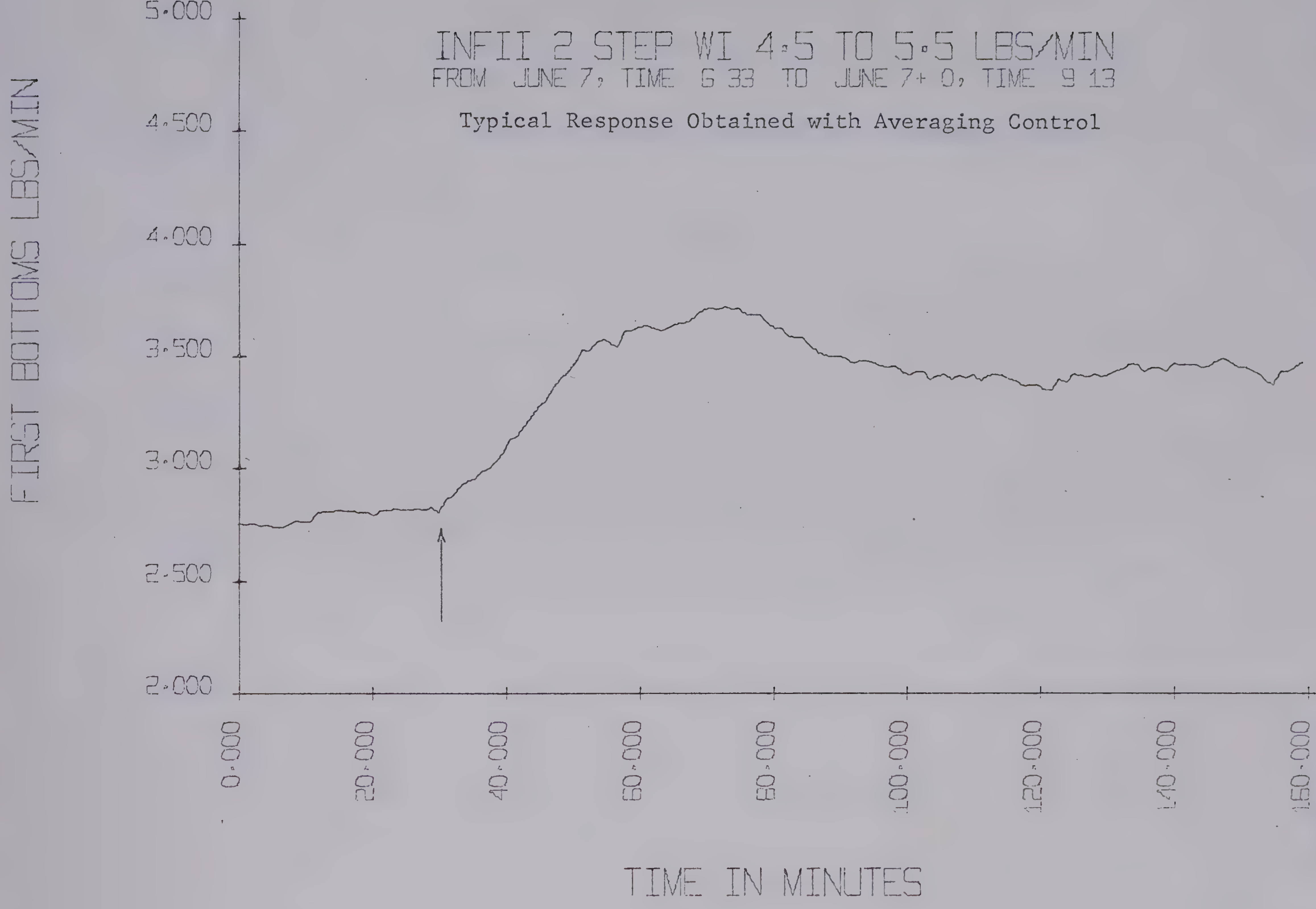

INFII 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 6 53 TO JUNE 7+ 0, TIME 9 1
Typical Response Obtained with Averaging Control
FIRST EFFECT LEVEL IN H2O
18.000
16.000
14.000
12.000
10.000
8.000
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
TIME IN MINUTES

Figure D-4



INFII 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 6 33 TO JUNE 7+ 0, TIME 9 13
Typical Response Obtained with Averaging Control
FIRST BOTTOMS LBS/MIN
5.000
4.500
4.000
3.500
3.000
2.500
2.000
0.000
20.000
40.000
60.000
80.000
100.000
120.000
140.000
160.000
TIME IN MINUTES

Figure D-5

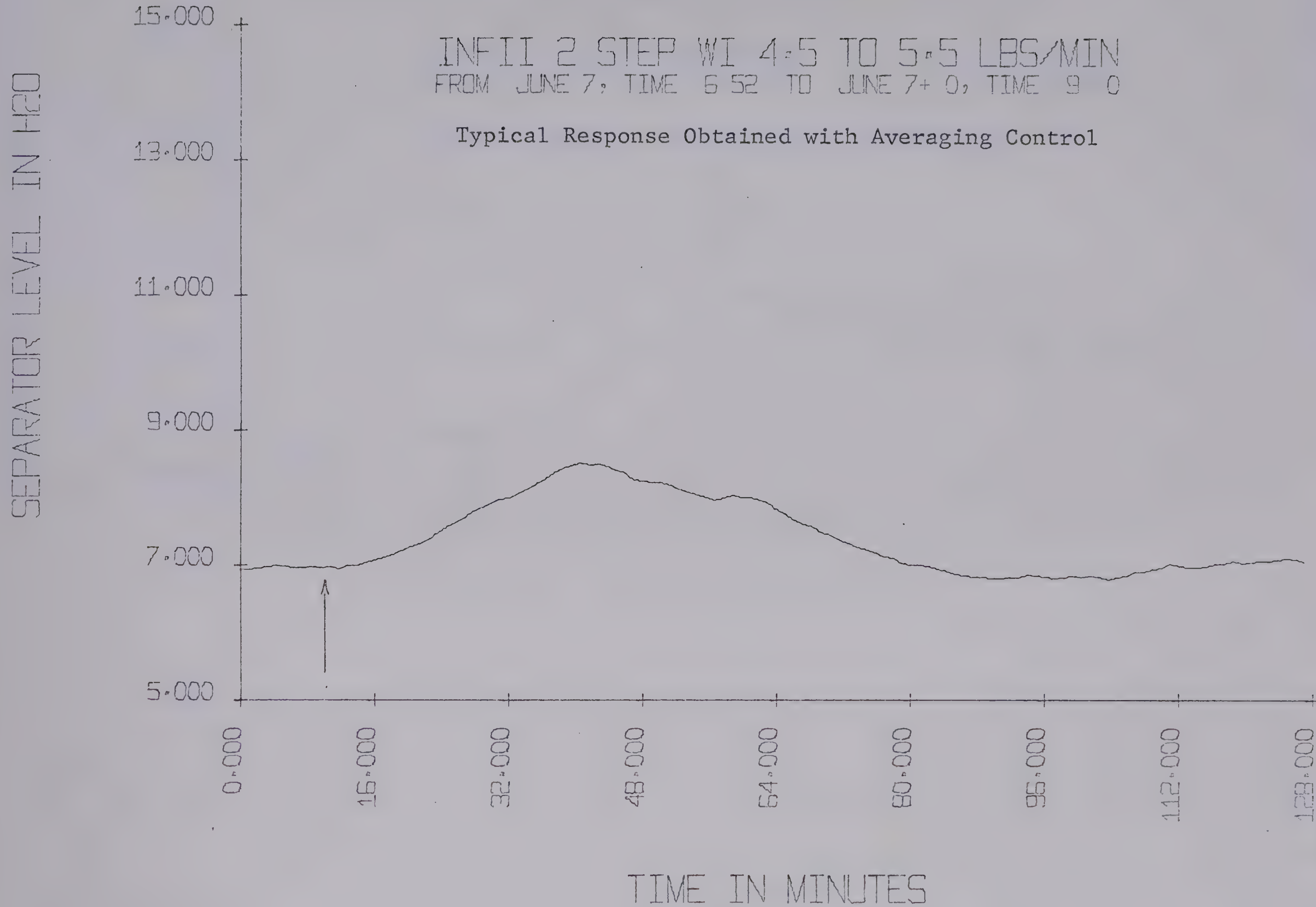

INFII 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 6 52 TO JUNE 7+ 0, TIME 9 0
Typical Response Obtained with Averaging Control
SEPARATOR LEVEL IN H2O
15.000
13.000
11.000
9.000
7.000
5.000
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
TIME IN MINUTES

Figure D-6

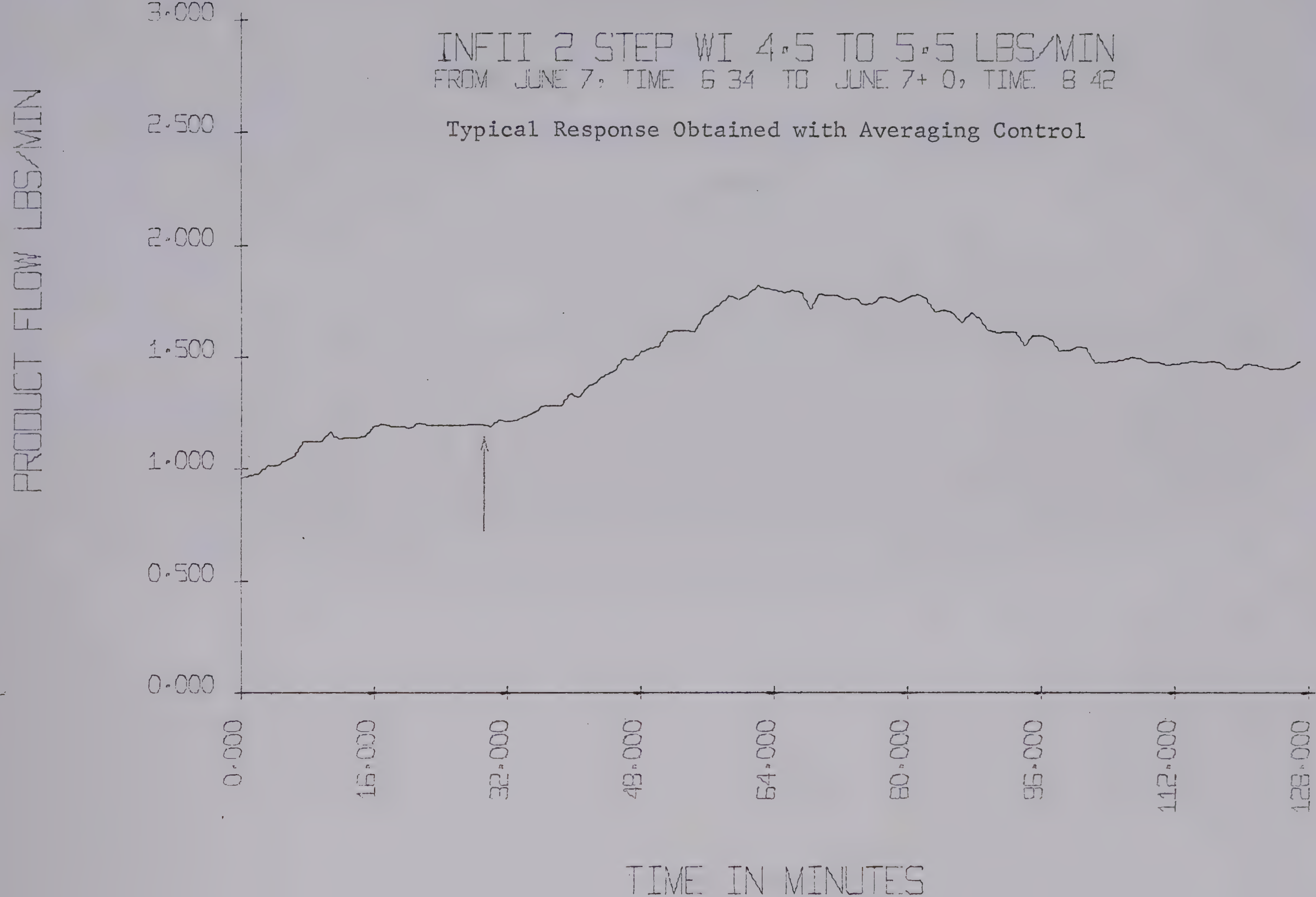
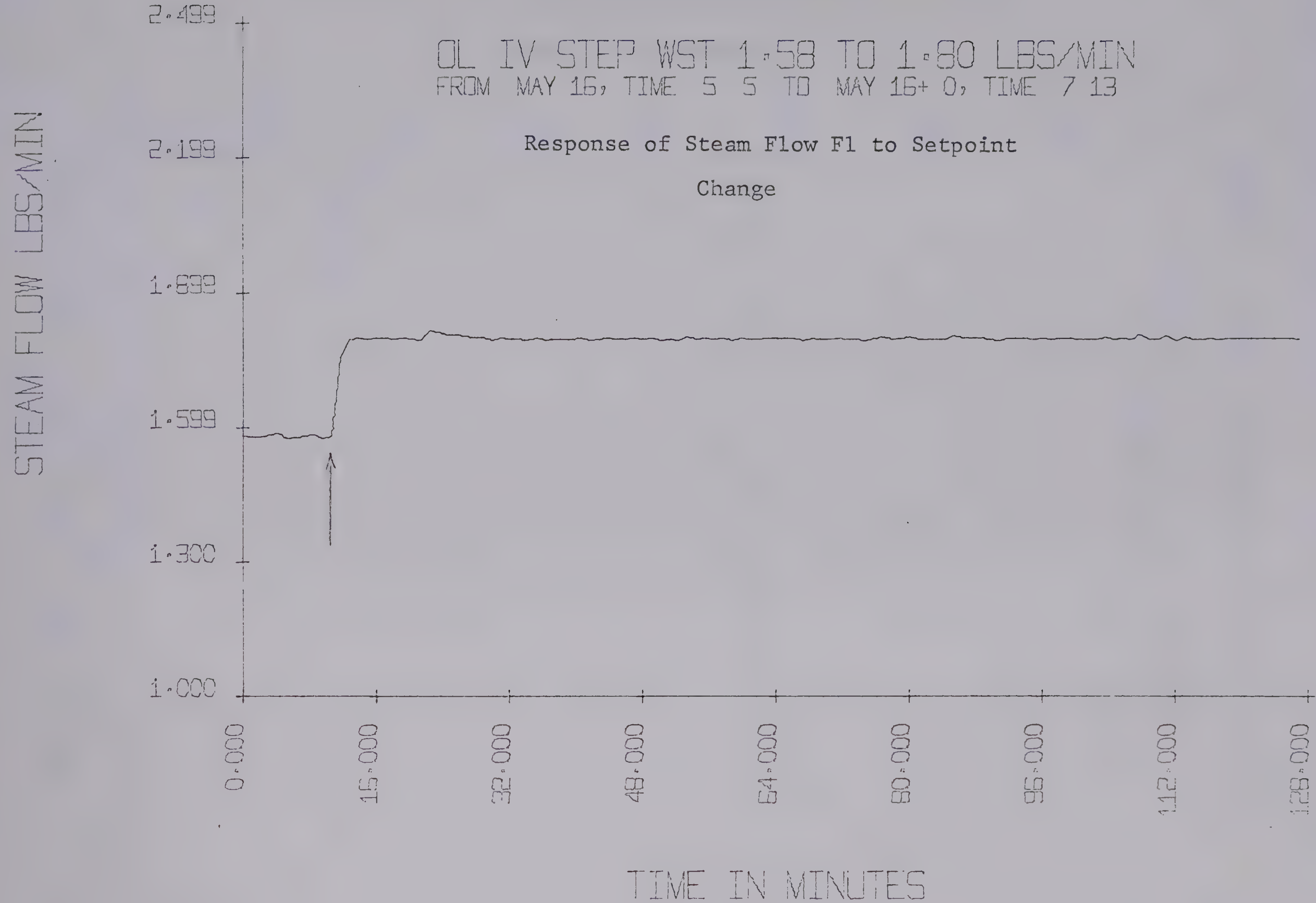

INFII 2 STEP WI 4.5 TO 5.5 LBS/MIN
FROM JUNE 7, TIME 6 34 TO JUNE 7+ 0, TIME 8 42
Typical Response Obtained with Averaging Control
PRODUCT FLOW LBS/MIN
3.000
2.500
2.000
1.500
1.000
0.500
0.000
0.000
16.000
32.000
48.000
64.000
80.000
96.000
112.000
128.000
TIME IN MINUTES

Figure D-7



Response of Steam Flow F1 to Setpoint Change

Figure D-8

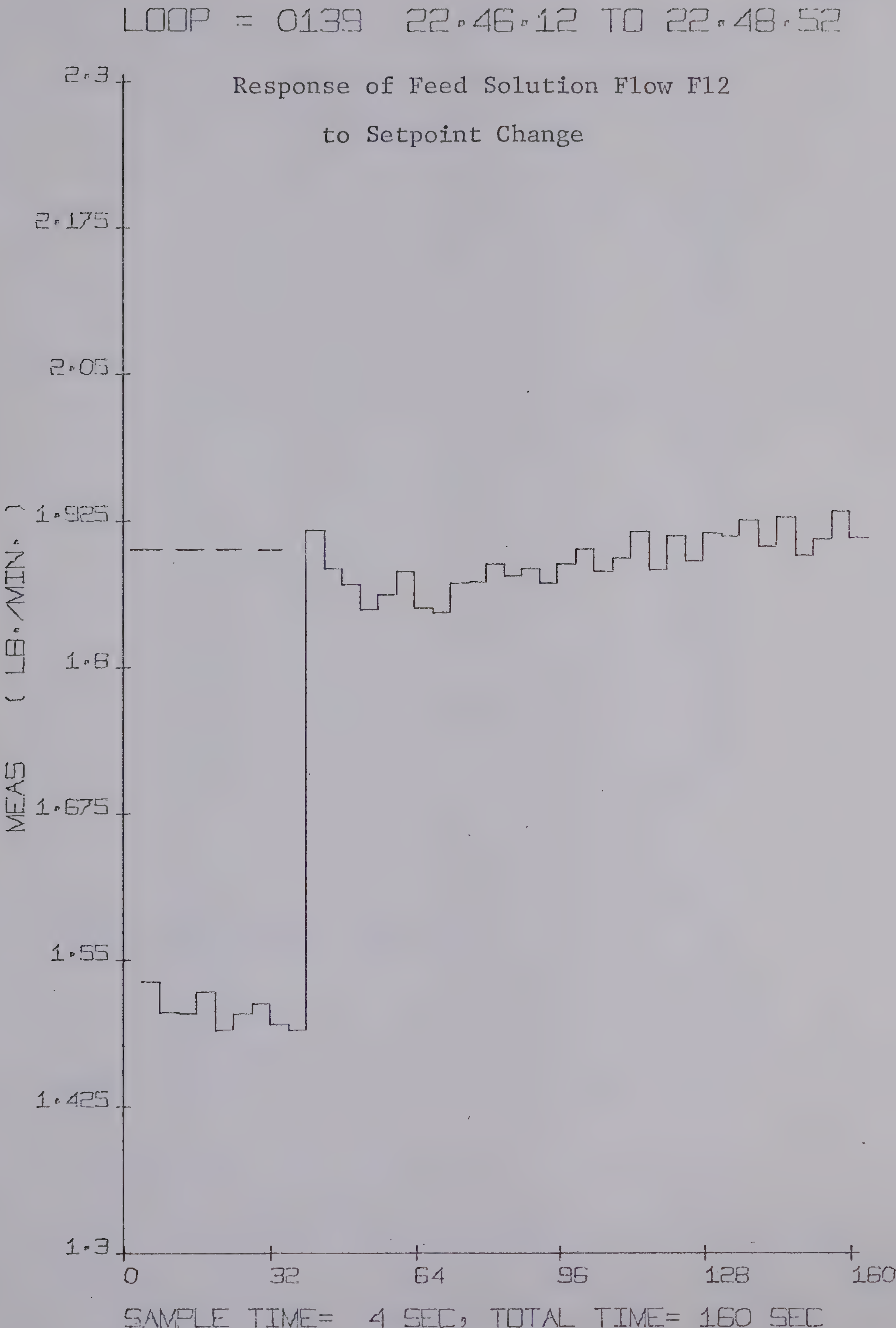

LOOP = 0139 22.46.12 TO 22.48.52
Response of Feed Solution Flow F12
to Setpoint Change
MEAS (LB./MIN.)
2.3
2.175
2.05
1.925
1.8
1.675
1.55
1.425
1.3
0
32
64
96
128
160
SAMPLE TIME= 4 SEC, TOTAL TIME= 160 SEC

Figure D-9

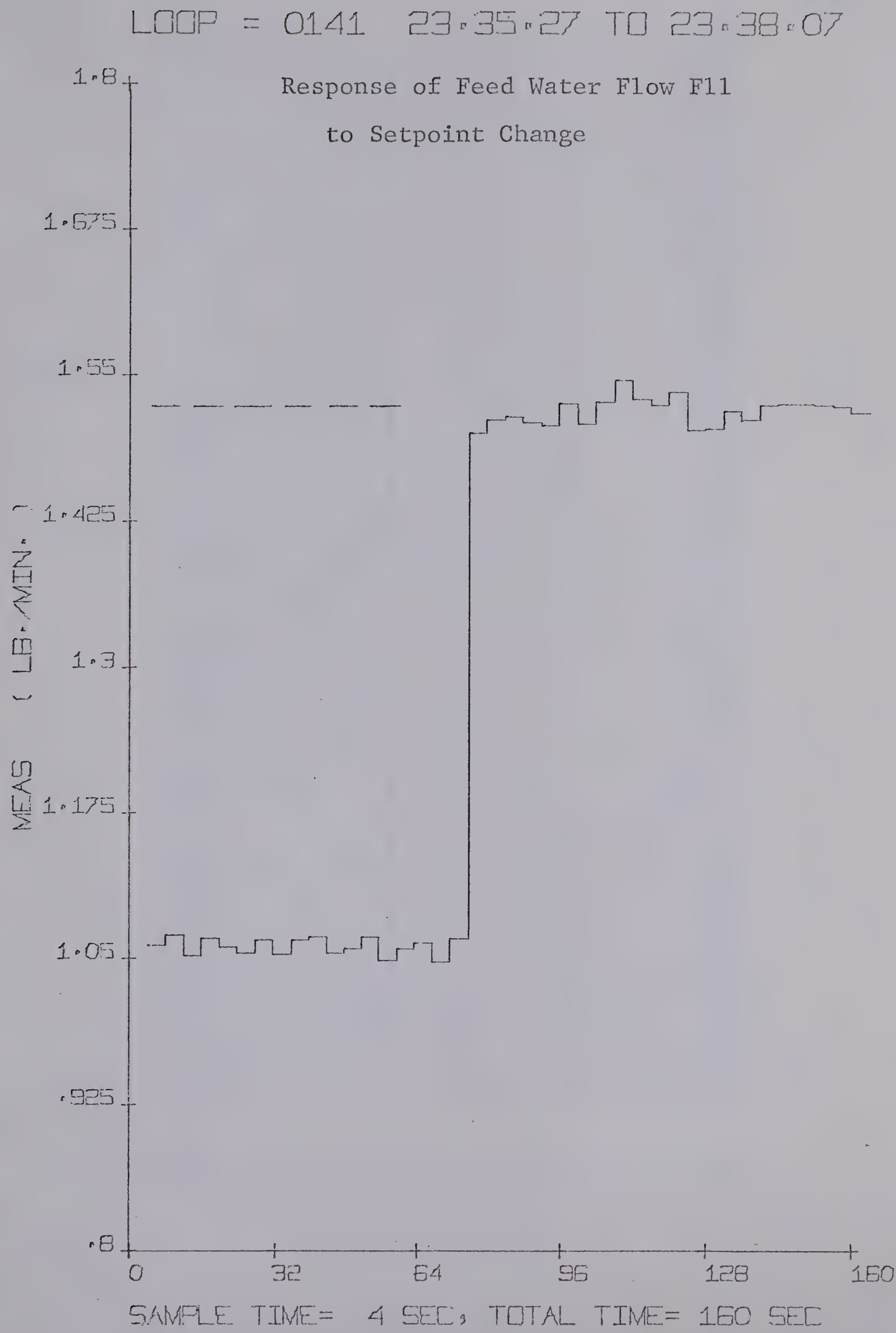
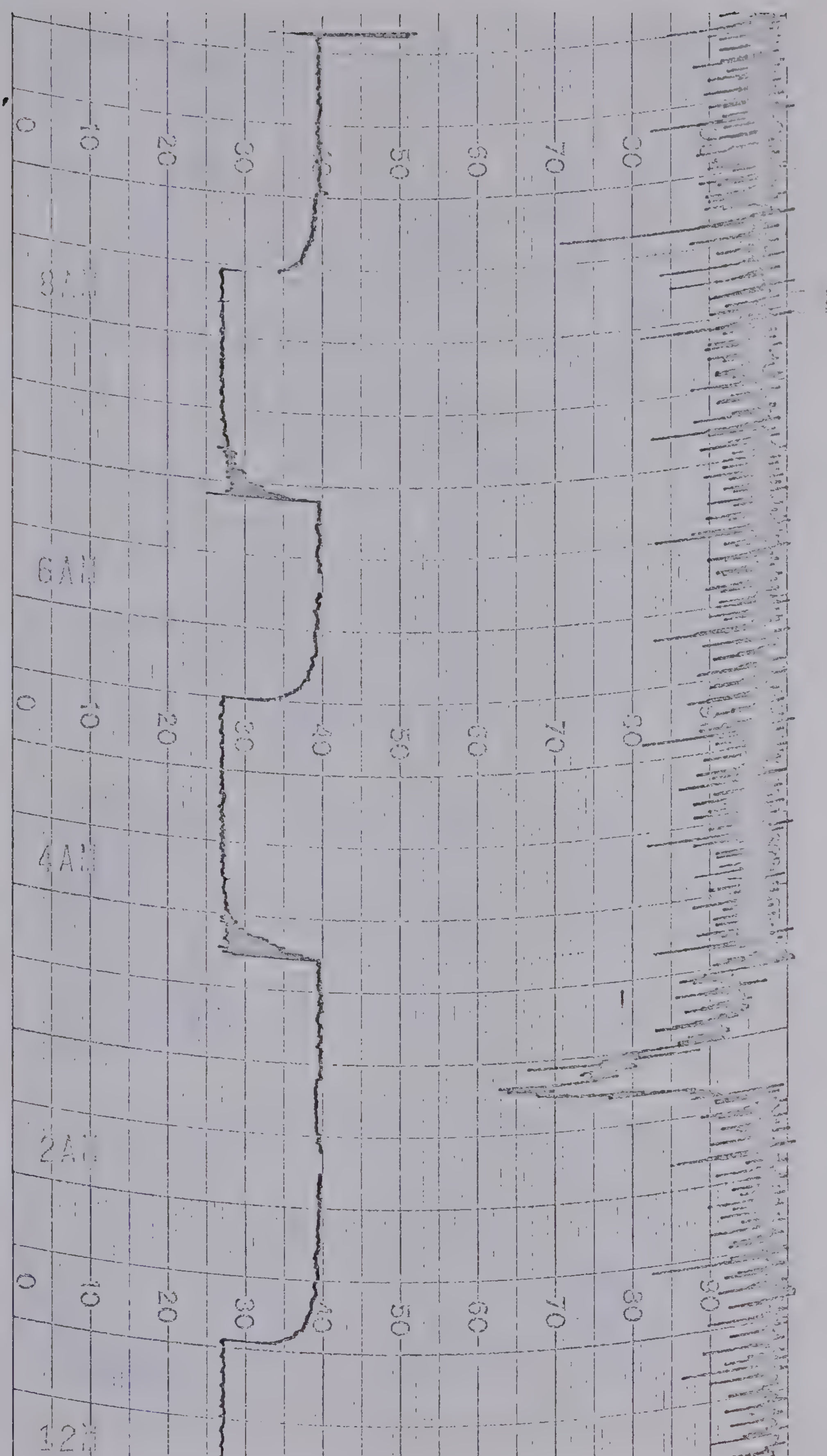

LOOP = 0141 23.35.27 TO 23.38.07
Response of Feed Water Flow F11
to Setpoint Change
MEAS (LB./MIN.)
1.8
1.675
1.55
1.425
1.3
1.175
1.05
.925
.8
0
32
64
96
128
160
SAMPLE TIME= 4 SEC, TOTAL TIME= 160 SEC

Figure D-10

F8

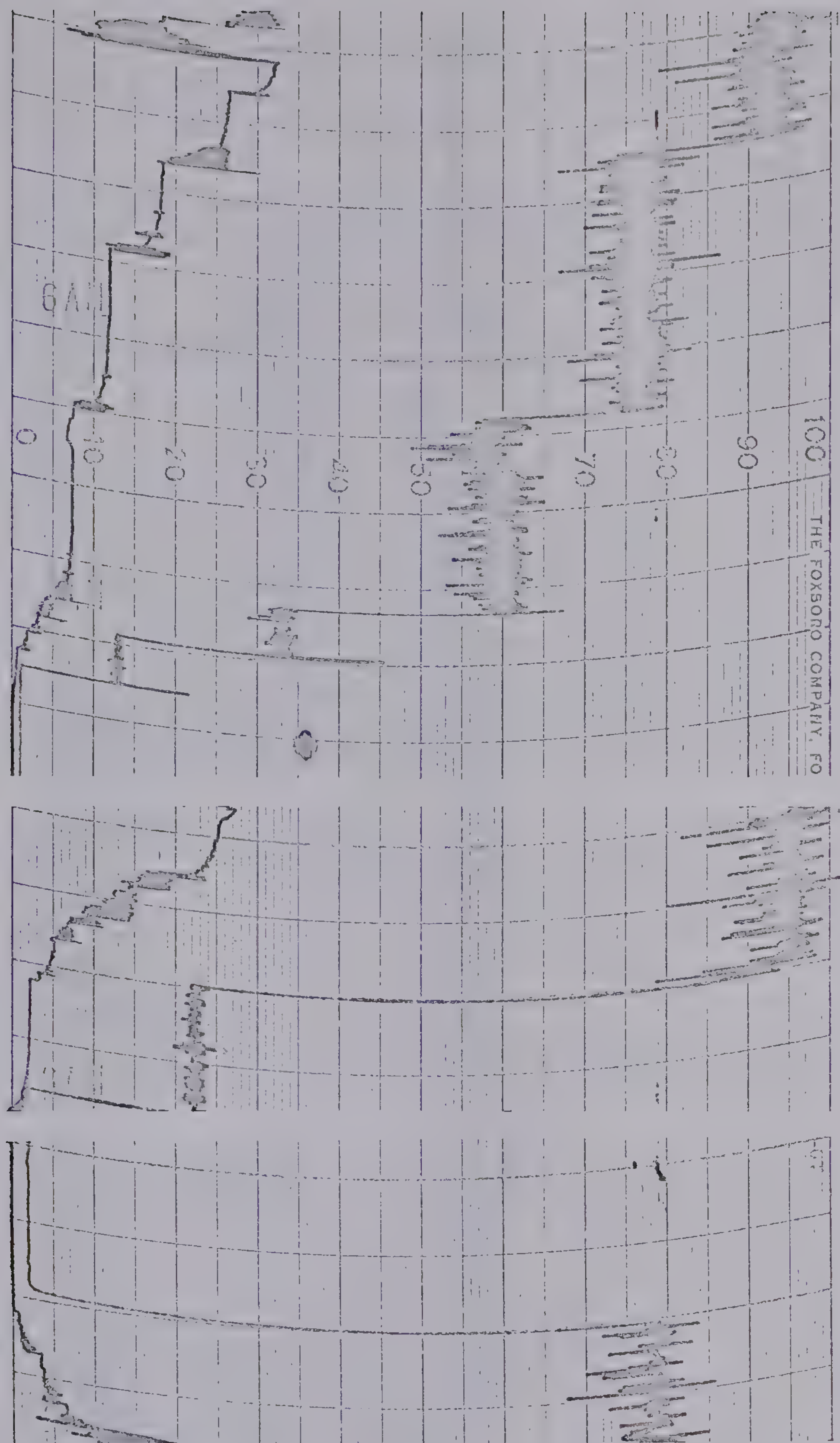


F9

Typical Responses of Total Feed Flow F8
to Setpoint Changes

Figure D-11

F8

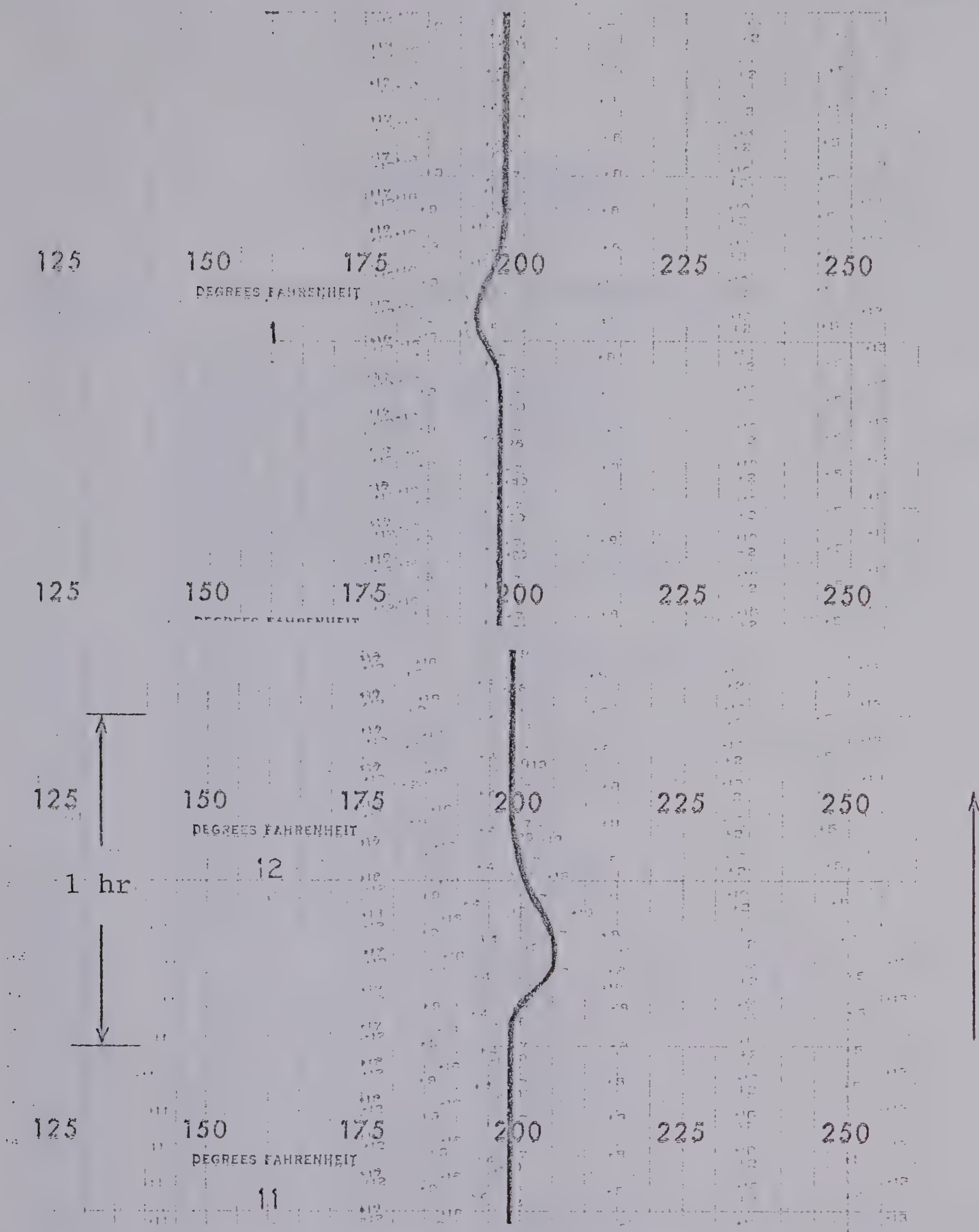


F9

Typical Responses of Cooling Water Flow F9

to Setpoint Changes

Figure D-12



Typical Responses of Feed Temperature T7 to a $\pm$ 1.0 lb/min Step Change in Feed Flow Rate

Figure D-13

POLLING SEQUENCE

TIME IN SECONDS	VARIABLE RECORDS POLLED
1	34 36 39 40 41 47 50 51 56
2	39 40 41 42 49 51 52 57 68
3	33 34 38 39 40 41 45 50 51 58
4	32 39 40 41 49 51 52 59 69
5	34 37 39 40 41 50 51 60
6	39 40 41 43 49 51 52 61
7	25 29 30 33 34 35 39 40 41 44 50 51 62
8	31 32 39 40 41 49 51 52 63 69
9	34 36 39 40 41 50 51 64
10	39 40 41 42 49 51 52 65 68
11	33 34 38 39 40 41 45 50 51 54 66
12	32 39 40 41 49 51 52 53 67 69
13	27 34 37 39 40 41 50 51
14	28 39 40 41 43 46 49 51 52
15	25 26 29 30 33 34 35 39 40 41 44 48 50 51
16	32 39 40 41 49 51 52 55 69
17	34 36 39 40 41 50 51 56
18	39 40 41 42 49 51 52 57 68
19	33 34 38 39 40 41 45 50 51 58
20	32 39 40 41 49 51 52 59 69
21	34 37 39 40 41 50 51 60
22	39 40 41 43 49 51 52 61
23	25 29 30 33 34 35 39 40 41 44 50 51 62
24	32 39 40 41 49 51 52 63 69
25	34 36 39 40 41 50 51 64
26	39 40 41 42 49 51 52 65 68
27	33 34 38 39 40 41 45 50 51 66
28	32 39 40 41 49 51 52 67 69
29	34 37 39 40 41 50 51
30	39 40 41 43 49 51 52
31	25 26 29 30 33 34 35 39 40 41 44 50 51
32	32 39 40 41 49 51 52 55 69

ETC.

TABLE D-4

POLLING SEQUENCE

TIME IN SECONDS	VARIABLE RECORDS POLLED
1	34 36 39 40 41 47 50 51 56
2	39 40 41 42 49 51 52 57 68
3	33 34 38 39 40 41 45 50 51 58
4	32 39 40 41 49 51 52 59 69
5	34 37 39 40 41 50 51 60
6	39 40 41 43 49 51 52 61
7	25 29 30 33 34 35 39 40 41 44 50 51 62.
8	31 32 39 40 41 49 51 52 63 69
9	34 36 39 40 41 50 51 64
10	39 40 41 42 49 51 52 65 68
11	33 34 38 39 40 41 45 50 51 54 66
12	32 39 40 41 49 51 52 53 67 69
13	27 34 37 39 40 41 50 51
14	28 39 40 41 43 46 49 51 52
15	25 26 29 30 33 34 35 39 40 41 44 48 50 51
16	32 39 40 41 49 51 52 55 69
17	34 36 39 40 41 50 51 56
18	39 40 41 42 49 51 52 57 68
19	33 34 38 39 40 41 45 50 51 58
20	32 39 40 41 49 51 52 59 69
21	34 37 39 40 41 50 51 60
22	39 40 41 43 49 51 52 61
23	25 29 30 33 34 35 39 40 41 44 50 51 62
24	32 39 40 41 49 51 52 63 69
25	34 36 39 40 41 50 51 64
26	39 40 41 42 49 51 52 65 69
27	33 34 38 39 40 41 45 50 51 66
28	32 39 40 41 49 51 52 67 69
29	34 37 39 40 41 50 51
30	39 40 41 43 49 51 52
31	25 26 29 30 33 34 35 39 40 41 44 50 51
32	32 39 40 41 49 51 52 55 69

ETC.

TABLE D-4

Appendix E

LIST OF DISK FILES USED

LOGICAL FILE NO.	FILE NAME	DATA STORED
20	DAFLE	DATA ACCUMULATION LOOP CONTENTS
21	JOSEF	SORTED VALUES FROM DAFLE FOR PLOTTING
25	BINGO	SUCCESSIVE MODEL OUTPUTS

LIST OF SKELETON COMMON WORDS USED

WORD NO.	VARIABLE NAME	VARIABLE DESCRIPTION
901-902	H	COMPUTATION INTERVAL IN SECONDS
903-904	CST2A	PROCESS MODEL CONSTANTS (SEE APP. C)
905-906	CST2W	
907-908	CST1B	
909-910	CST1C	
911-912	CST9C	
913-914	CST9W	
915-916	CST5D	
917-918	CST5W	
919-920	CST4E	
921-922	CST4C	
923-924	CSTTF	
925-926	CSTTW	
927-928	A	INTERMEDIATE QUANTITIES ON FIGURE 5.4
929-930	B	
931-932	C	
933-934	D	
935-936	E	
937-938	F	
939-940	STONE	DUMMY VARIABLES FOR LATER ADDITIONS
941-942	BRICK	
943-944	WALL	
945-946	ERICA	REAL TIME DIFFERENCE BETWEEN FIRST VALUES IN DAFLE AND BINGO IN MINUTES
947	IRMA	FLAG 6 ON FIGURE 5.5
948	LOLA	SECTOR ADDRESS IN FILE BINGO
949	JOSEY	FLAG 5 ON FIGURE 5.5
950-951	CI	PRESENT VALUE OF FEED CONCENTRATION
952-953	WI	PRESENT VALUE OF FEED FLOW RATE
954-955	WST	PRESENT VALUE OF STEAM FLOW RATE
956	INES	TIME ELAPSED (NUMBER OF 64 SEC. INTERVALS)BEFORE CONTROLLER USES MODEL OUTPUT

TABLE E-1

MAINLINE BEGIN

```
      MAINLINE BEGIN

C
C          PROGRAM NAME        BEGIN
C
C          PROGRAM TYPE        PROCESS PROGRAM, MAINLINE CORE
C                              LOAD
C
C          ABSTRACT            THE PROGRAM SELECTIVELY ACTIVATES
C                              OR DEACTIVATES DDC LOOP RECORDS
C                              OR INITIALIZES DATA TRANSFER
C                              FROM RING BUFFERS TO DISK
C
C          OPERATION           PROGRAM IS CONTACTED VIA FUNCTION
C                              NO. 5 OR KEYBOARD REQUEST
C
C          SUBROUTINES         FFINP, DAINL, RSBUF, OPER, NONOP
C          CALLED              VIAQ
C

      DATA JTYPE,JREAD/2,1/
      DEFINE FILE 20(40,320,U,KK)

C          DECIDE ON WHICH FUNCTION BEGIN IS TO PERFORM
C

      WRITE(JTYPE,101)
 101  FORMAT(//T5,'ENTER 1 TO ACTIVATE EVAPORATOR LOOPS',/T5
     1,'ENTER 2 T
     1 DEACTIVATE EVAPORATOR LOOPS',/T5,'ENTER 3 TO ACTIVATE
     1 DISK FILE'
 110  CALL FFINP(JREAD,1,0,I,IEROR)
      IF(IEROR) 112,116,112
 112  WRITE(JTYPE,1121)
 1121 FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 110
 116  IF(I) 112,112,121
 121  IF(I-3) 126,126,112
 126  GO TO (1265,140,150), I

C          BEGIN IS TO ACTIVATE LOOPS
C          OBTAIN ID OF FIRST AND LAST LOOPS IN SEQUENCE
C

 1265 WRITE(JTYPE,127)
 127  FORMAT(T5,'ENTER ID NUMBERS OF FIRST AND LAST LOOPS IN
     1 SEQUENCE T
     1 BE ACTIVATED')
 128  CALL FFINP(JREAD,2,0,IDF,0,IDL,IEROR)
      IF(IEROR) 1281,1285,1281
```

MAINLINE BEGIN

```
 1281 WRITE(JTYPE,1121)
      GO TO 128
 1285 CONTINUE
      DO 129  J=IDF,IDL
      CALL OPER(J)
 129  CONTINUE
      WRITE(JTYPE,1291)
 1291 FORMAT(T5,'LOOP ACTIVATION COMPLETED')
      CALL VIAQ

C         BEGIN IS TO DEACTIVATE LOOPS
C

 140  WRITE(JTYPE,141)
 141  FORMAT(T5,'ENTER ID NUMBERS OF FIRST AND LAST LOOPS IN
     1 SEQUENCE T
     1 BE DEACTIVATED')
 142  CALL FFINP(JREAD,2,0,IDF,0,IDL,IEROR)
      IF(IEROR) 143,144,143
 143  WRITE(JTYPE,1121)
      GO TO 142
 144  CONTINUE
      DO 145  J=IDF,IDL
      CALL NONOP(J)
 145  CONTINUE
      WRITE(JTYPE,146)
 146  FORMAT(T5,'LOOP DEACTIVATION COMPLETED')
      CALL VIAQ

C         BEGIN IS TO ACTIVATE DISK FILE
C

 150  CALL RSBUF
      CALL DAINL
      WRITE(JTYPE,151)
 151  FORMAT(T5,'DATA TRANSFER TO DISK HAS BEEN
     1 INITIALIZED')
      CALL VIAQ
      END
```

MAINLINE THIS

MAINLINE THIS

```
C
C        PROGRAM NAME        THIS
C
C        PROGRAM TYPE        PROCESS PROGRAM. MAINLINE CORE
C                            LOAD
C
C        ABSTRACT            THE PROGRAM PRODUCES A GRAPH OF
C                            VALUES COLLECTED IN THE DISK
C                            FILE VERSUS TIME
C
C        OPERATION           THE PROGRAM IS CONTACTED VIA
C                            FUNCTION NO. 5 OR KEYBOARD
C                            REQUEST
C
C        SUBROUTINES CALLED  FFINP, GTDDL, GBDAT, SCALF
C                            FCHAR, MDGP, FPLOT, VIAQ
C
C

      DIMENSION IFR(3), ICOM(7), ITO(3)
      DIMENSION VEC(30), IHEAD(20), LABX(16), EDIT(16)
      DIMENSION IFUNC(4), IFCB1(323), IFCB2(33), IVEC(30)
      DIMENSION ABMIN(30), LABY(32), ILA(16)
      DATA ITO,IPLUS/'  ','TO','  ','+'/
      DATA IFR,ICOM/'FR','OM','  ',',',' ','T','I','M','E','
     1 '/
      DATA JTYPE,JREAD,KPLOT,ICHAR,NAXIS,ISORT/2,1,2,0,1,1/
      DATA XMIN,NN/0.0,30/
      DATA LABX/'T','I','M','E',' ','I','N',' ','M','I','N'
     1,'U','T','E'
     1'S','''/
      DATA EDIT/1000.0,10000.0,100000.0,1000000.0,1.0,10.0
     1,100.0,1000.0
     10.001,0.01,0.1,1.0,4*0.0001/
      DEFINE FILE 20(40,320,U,KK), 21(50,30,U,LL)
      IFUNC(2)=1500
      IFCB1(1)=20
      IFCB1(2)=320
      IFCB1(3)=40
      IFCB2(1)=21
      IFCB2(2)=30
      IFCB2(3)=50
      WRITE(JTYPE,1401)
 1401 FORMAT(T5,'ENTER ID OF DATA ACQUISITION LOOP, POLL
     1 TIME INTERVAL
     1N SECONDS'/T5,'AND DATE AND TIME AT WHICH THIS LOOP
     1 WAS MADE OPER
     1BLE'/T5,'USING THE FORMAT OF THE FOLLOWING EXAMPLE,
```

MAINLINE THIS

```
     1 0148,128,JA
     1 ,07,14,06')
 145  CALL FFINP(JREAD,6,0,LPID,0,LPLT,3,RMON1,0,IDAT1,0
     1,IHR1,0,IMIN1,I
     1ROR)
      IF(IEROR) 146,147,146
 146  WRITE(JTYPE,1461)
 1461 FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 145
C
C        CONVERT THE HEXADECIMAL LOOP ID INTO A DECIMAL
C        NUMBER
C
 147  LPID1=LPID/100
      LPID3=LPID-100*LPID1
      LPID2=LPID3/10
      LPID3=LPID3-10*LPID2
      LPID4=LPID1*256+LPID2*16+LPID3
C
C        LPID4 CONTAINS THE DATA ACQUISITION LOOP ID IN FORM
C        OF A
C        DECIMAL NUMBER
C        BEGIN DATA TRANSFER FROM DAFLE TO TEMPORARY FILE
C        NO. 21
C
      IFUNC(1)=LPID4
      CALL GBDAT(IFUNC,IFCB1,IFCB2)
C
C        VERIFY SUCCESS OF DATA TRANSFER
C
      JINT=IFUNC(3)
      GO TO (19,15,16,17,18), JINT
 15   WRITE(JTYPE,151) LPID
 151  FORMAT(T5,'LOOP',T11,I4,T15,'NOT FOUND IN DISK FILE'
     1/T5,'IF YOU W
     1NT TO TRY AGAIN, CALL FUNCTION NO. 5')
      GO TO 28
 16   WRITE(JTYPE,161) IFUNC(2)
 161  FORMAT(T5,I4,T11,'VALUES HAVE BEEN FOUND IN DAFLE AND'
     1/T5,'HAVE B
     1EN TRANSFERRED TO THE TEMPORARY FILE')
      GO TO 20
 17   WRITE(JTYPE,171)
 171  FORMAT(T5,'THE DISK FILE IS EMPTY AT PRESENT, TRY
     1 LATER')
      GO TO 28
 18   WRITE(JTYPE,181)
```

MAINLINE THIS

```
 181  FORMAT(T5,'RECORD SIZE OF INPUT FILE IFCB1 IS
     1 NEGATIVE'/T5,'TRANS
     1ER INTERRUPTED, YOU ARE OFF')
      GO TO 28
 19   WRITE(JTYPE,191) IFUNC(2)
 191  FORMAT(T5,'DATA TRANSFER TO TEMPORARY FILE COMPLETED'
     1/T5,'NUMBER
     1F POINTS TRANSFERRED IS',T37,I4)

C        DATA ARE READY TO BE ACCESSED IN TEMPORARY FILE.
C        START PREPARING THE GRAPH.
C        CALCULATE DATE AND TIME OF LAST POINT IN TEMPORARY
C        FILE.
C

 20    ITOSE=LPLT*IFUNC(2)
       ITOMI=ITOSE/60
       ITOHR=ITOMI/60
       ITODA=ITOHR/24
       IREMI=ITOMI-ITOHR*60
       IREHR=ITOHR-ITODA*24
       IFIHR=IHR1+IREHR
       IFIMI=IMIN1+IREMI
       IF(60-IFIMI) 202,202,203
 202   IFIMI=IFIMI-60
       IFIHR=IFIHR+1
 203   IF(24-IFIHR) 204,204,205
 204   IFIHR=IFIHR-24
       ITODA=ITODA+1

C        THE TIME AT WHICH THE LAST POINT WAS STORED IN
C        DAFLE IN DAYS,
C        HOURS, AND MINUTES IS CONTAINED IN ITODA, IFIHR,
C        AND IFIMI,
C        RESPECTIVELY
C
C        DATA WILL BE MOVED OUT OF THE TEMPORARY FILE AT THE
C        RATE OF ONE
C        RECORD (30 POINTS) AT A TIME.
C        OBTAIN INPUT PARAMETERS FOR MDGP.
C

 205   TOMIN=FLOAT(ITOSE)/60.0
       WRITE(JTYPE,206) TOMIN
 206   FORMAT(T5,'TOTAL TIME TO APPEAR ON ABSCISSA IS',T41
     1,F7.1,T50,'MIN
     1TES')
       WRITE(JTYPE,210)
 210   FORMAT(T5,'ENTER INFORMATION ON GRAPH IN THIS ORDER'
     1/T5,'MAXIMUM
```

```
     1ALUE OF DEPENDENT VARIABLE(REAL)'/T5,'MINIMUM VALUE OF
     1 DEPENDENT
     1ARIABLE(REAL)'/T5,'NUMBER OF DIVISIONS ALONG
     1 ABSCISSA(INTEGER)'/T
     1,'NUMBER OF DIVISIONS ALONG ORDINATE(INTEGER)')
 211  CALL FFINP(JREAD,4,1,YMAX,1,YMIN,0,NX,0,NY,IEROR)
      IF(IEROR) 212,213,212
 212  WRITE(JTYPE,2121)
 2121 FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 211
 213  WRITE(JTYPE,2131)
 2131 FORMAT(T5,'ENTER LABEL FOR ORDINATE (32 ELEMENTS AT
     1 MOST, ENDING
     1ITH '' , LEFT JUSTIFIED)')
 214  CALL FFINP(JREAD,1,162,ILA(1),IEROR)
      IF(IEROR) 215,216,215
 215  WRITE(JTYPE,2121)
      GO TO 214

C          CONVERT LABEL FROM A2 TO A1 FORMAT
C

 216  DO 218 L=1,16
      M=L*2-1
      K1=ILA(L)/4096
      K2=ILA(L)-K1*4096
      K3=K2/256
      IF(K3)2162,2163,2163
 2162 LABY(M)=K1*4096+(K3-1)*256+64
      GO TO 2165
 2163 LABY(M)=K1*4096+K3*256+64
 2165 M=L*2
      K4=ILA(L)-K1*4096-K3*256
      K5=K4/16
      K6=K4-K5*16
      LABY(M)=K5*4096+K6*256+64
 218  CONTINUE

C          DEFINE FIRST 30 ELEMENTS OF TIME VECTOR
C

      IABSC=0
      ABMIN(1)=0.0
      DO 22  M=2,NN
      IABSC=IABSC+LPLT
      ABMIN(M)=FLOAT(IABSC)/60.0
 22   CONTINUE

C          DEFINE FIRST 30 ELEMENTS OF DEPENDENT VARIABLE
C          VECTOR.
```

MAINLINE THIS

```
C         CONVERT DATA IN TEMPORARY DISK FILE TO ENGINEERING
C         UNITS.
C

      CALL GTDDL(LPID4,IDSPY,ICONA,ICONB,IEDIT,IUNIT,IER)
      IF(IER) 223,222,223
 222  WRITE(JTYPE,2221)
 2221 FORMAT(T5,'DONOR LOOP COULD NOT BE FOUND, DATA
     1 TRANSFER INTERRUPT
     1D'/T5,'IF YOU LIKE TO START AGAIN, CALL FUNCTION NO.
     1 5')
      GO TO 28
 223  CONB=FLOAT(ICONB)*EDIT(IEDIT)
      CONA=FLOAT(ICONA)*EDIT(IEDIT)/65536.0
      K=1
      READ(21'K) (IVEC(L),L=1,NN)
      DO 24  MM=1,NN
      VEC(MM)=CONA*FLOAT(IVEC(MM))+CONB
 24   CONTINUE

C         START GRAPH, PLOTTING FIRST 30 POINTS.
C

      WRITE(JTYPE,245)
 245  FORMAT(T5,'ENTER TITLE OF PLOT AS A VECTOR OF 40
     1 CHARACTERS')
 2452 CALL FFINP(JREAD,1,202,IHEAD,IEROR)
      IF(IEROR) 246,247,246
 246  WRITE(JTYPE,2121)
      GO TO 2452
 247  CALL SCALF(1.0,1.0,0.0,0.0)
      CALL FCHAR(3.5,7.2,0.15,0.2,0.0)
      WRITE(7,248)IHEAD
 248  FORMAT(20A2)
      CALL FCHAR(3.95,7.0,0.1,0.13,0.0)
      WRITE(7,249) IFR, RMON1, IDAT1, ICOM, IHR1, IMIN1, ITO
     1, RMON1, ID
     1T1, IPLUS, ITODA, ICOM, IFIHR, IFIMI
 249  FORMAT(3A2,A4,I2,7A1,I2,I3,3A2,A4,I2,A1,I2,7A1,I2,I3)
      CALL FPLOT(+5,0.0,0.0)
      CALL MDGP(ABMIN,VEC,NN,TOMIN,XMIN,YMAX,YMIN,NX,NY
     1,KPLOT,ICHAR,NAX
     1S,ISORT,LABX,LABY)

C         PLOT THE REMAINDER OF THE DATA IN THE TEMPORARY
C         FILE ON THE
C         SAME GRAPH.
C

 262  JRE=IFUNC(2)-K*NN
```

MAINLINE THIS

```
      IF(JRE-NN) 263,264,264
263   NN=JRE
264   DO 265  KL=1,NN
      IABSC=IABSC+LPLT
      ABMIN(KL)=FLOAT(IABSC)/60.0
265   CONTINUE
      K=K+1
      READ(21'K) (IVEC(L),L=1,NN)
      DO 266  KM=1,NN
      VEC(KM)=CONA*FLOAT(IVEC(KM))+CONB
266   CONTINUE
      DO 267  KP=1,NN
      CALL FPLOT(0,ABMIN(KP),VEC(KP))
267   CONTINUE
      IF(JRE-NN) 268,268,262
268   XNEW=TOMIN+4.0*(TOMIN-XMIN)/8.0
      YNEW=YMIN-2.5*(YMAX-YMIN)/5.0
      CALL FPLOT(+5,XNEW,YNEW)
      WRITE(JTYPE,270)
270   FORMAT(T5,'LIKE IT OR NOT, THIS IS IT')
28    CONTINUE
      CALL EXIT
      END
```

MAINLINE FDFWD

```
      MAINLINE FDFWD
C
C        PROGRAM NAME             FDFWD
C
C        PROGRAM TYPE             PROCESS PROGRAM, MAINLINE
C                                 CORE LOAD
C
C        ABSTRACT                 THIS PROGRAM PERFORMS
C                                 MATERIAL BALANCES AROUND
C                                 VARIOUS SECTIONS OF THE
C                                 EVAPORATOR BASED ON FLOW AND
C                                 CONCENTRATION READINGS
C                                 OBTAINED FROM THE
C                                 PROCESS VARIABLE TABLE
C
C        OPERATION                PROGRAM IS CONTACTED VIA
C                                 FUNCTION NO. 5 OR
C                                 KEYBOARD REQUEST
C
C        SUBROUTINES CALLED       JACK, QUEEN, KING, GTVLU
C                                 VIAQ
C
C
      CALL JACK(JUDY)
      GO TO (2,3), JUDY
 2    CALL QUEEN
      CALL KING
 3    CALL VIAQ
      END
```

SUBROUTINE JACK

```
      SUBROUTINE JACK(JUDY)
C        THIS SUBROUTINE TAKES READINGS REQUIRED FOR
C        MATERIAL BALANCE
C        FROM PROCESS VARIABLE TABLE
C
      COMMON F1,F2,F5,F6,F7,F8,CI,CO
      JUDY=1
      CALL GTVLU(306,1,F1,LULU,4)
      GO TO (2,9,9,9), LULU
 2    CALL GTVLU(337,1,F2,LULU,4)
      GO TO (3,9,9,9), LULU
 3    CALL GTVLU(336,1,F5,LULU,4)
      GO TO (4,9,9,9), LULU
 4    CALL GTVLU(338,1,F6,LULU,4)
      GO TO (5,9,9,9), LULU
 5    CALL GTVLU(294,1,F7,LULU,4)
      GO TO (55,9,9,9), LULU
 55   CALL GTVLU(329,1,F8,LULU,4)
      GO TO (6,9,9,9), LULU
 6    CALL GTVLU(323,1,CI,LULU,1)
      GO TO (7,9,9,9), LULU
 7    CALL GTVLU (324,1,CO,LULU,1)
      GO TO (95,9,9,9), LULU
 9    WRITE(2,91) LULU
 91   FORMAT(T5,'GTVLU ERROR CODE',I3,T25,'EXECUTION
     1 INTERRUPTED')
      JUDY=2
 95   RETURN
      END
```

SUBROUTINE QUEEN

```
      SUBROUTINE QUEEN
C
C         THIS SUBROUTINE CALCULATES ERRORS OF CLOSURE FOR
C         PRESENT
C         STEADY STATE CONDITIONS
C
      COMMON F1,F2,F5,F6,F7,F8,CI,CO,STEC,E1,E2,E3,E4,E5
      STEC=(F8-F8*CI/CO)/F1
      E1=(F8-F5-F7-F6)*100.0/F8
      E2=(F8-F5-F2)*100.0/F8
      E3=(F2-F6-F7)*100.0/F2
      E4=(F8*CI-F6*CO)*100.0/(F8*CI)
      E5=(F8*(100.0-CI)-F6*(100.0-CO)-F1*STEC*100.0)/(F1
     1*STEC)
      RETURN
      END
```

SUBROUTINE KING

```
      SUBROUTINE KING
C
C        THIS SUBROUTINE WRITES ALL ERRORS OF CLOSURE ON THE
C        TYPEWRITER
C
      DIMENSION FAYE(8)
      COMMON FAYE,STEC,E1,E2,E3,E4,E5
      WRITE(2,10) STEC
 10   FORMAT(//T5,'PRESENT STEAM ECONOMY IS',F8.3)
      WRITE(2,11)
 11   FORMAT(/T5,'PRESENT ERRORS OF CLOSURE IN PER CENT
     1 ARE')
      WRITE(2,21) E1
 21   FORMAT(T5,'TOTAL PROCESS',F8.1)
      WRITE(2,31) E2
 31   FORMAT(T5,'FIRST EFFECT',F9.1)
      WRITE(2,41) E3
 41   FORMAT(T5,'SECOND EFFECT',F8.1)
      WRITE(2,51) E4
 51   FORMAT(T5,'SOLUTE',T18,F8.1)
      WRITE(2,61) E5
 61   FORMAT(T5,'OVERHEAD FLOWS',F7.1)
      WRITE(2,71)
 71   FORMAT(T5,'END'//)
      RETURN
      END
```

MAINLINE INFER

```
      MAINLINE INFER

C
C        PROGRAM NAME              INFER
C
C        PROGRAM TYPE              PROCESS PROGRAM, MAINLINE
C                                  CORE LOAD
C
C        ABSTRACT                  PROGRAM SUPERVISES
C                                  INFERENTIAL CONTROL BY
C                                  PROVIDING BRANCHES TO VARIOUS
C                                  CORE LOAD
C
C        OPERATION                 PROGRAM IS CONTACTED VIA
C                                  FUNCTION NO. 5 OR
C                                  KEYBOARD REQUEST
C
C        SUBROUTINES CALLED        ZERO, VIAQ, CHAIN, CANCL
C                                  FFINP
C
C

      EXTERNAL EDITH, DIANE
      CALL ZERO(JANE)
      GO TO (10,20,30,40), JANE
 10   CALL CHAIN(DIANE)
 20   CALL CHAIN(EDITH)
 30   CALL CANCL(4)
 40   CALL VIAQ
      END
```

SUBROUTINE ZERO

```
      SUBROUTINE ZERO(JANE)
      WRITE(2,50)
   50 FORMAT(T5,'THESE CHOICES ARE OPEN TO YOU NOW'/T5,' 1
     1 START COLLE
     1TING INFORMATION FOR INFERENTIAL CONTROLLER'/T5,' 2
     1 GET TIMER NO
     2 4 ON ITS WAY'/T5,' 3  CEASE INFERENTIAL CONTROL'/T5,'
     1 4  CALL IT
     3A DAY')
   60 CALL FFINP(1,1,0,JANE,IRENE)
      IF(IRENE) 70,80,70
   70 WRITE(2,71)
   71 FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 60
   80 IF(JANE) 70,70,90
   90 IF(JANE-4) 120,100,70
  100 WRITE(2,101)
  101 FORMAT(T5,'OKAY, YOU ARE OFF')
  120 RETURN
      END
```

MAINLINE DIANE

```
      MAINLINE DIANE

C
C        PROGRAM NAME           DIANE
C
C        PROGRAM TYPE           PROCESS PROGRAM, MAINLINE
C                               CORE LOAD
C
C        ABSTRACT               PROGRAM PROVIDES ALL CON-
C                               STANTS AND PARAMETERS
C                               FOR INFERENTIAL CONTROLLER
C
C        OPERATION              PROGRAM IS CONTACTED VIA CALL
C                               CHAIN FROM INFER
C
C        SUBROUTINES CALLED     ONE, TWO, THREE, FOUR, SIX
C                               CHAIN, FFINP, VIAQ, GTVLU
C
C
      EXTERNAL EDITH
      DIMENSION GAIL(450)
      COMMON/INSKEL/GAIL,H,CST2A,CST2W,CST1B,CST1C,CST9C
     1,CST9W,CST5D
      COMMON/INSKEL/CST5W,CST4E,CST4C,CSTTF,CSTTW,A,B,C,D,E
     1,F
      COMMON/INSKEL/STONE,BRICK,WALL
      COMMON HOLD1,HOLD2,STEC,CI,CO,WI,WST,C2,FLOOR,SKY,DUST
 10   CALL ONE
      CALL TWO(JULIA)
      GO TO (20,90), JULIA
 20   CALL THREE(JEAN)
      GO TO (10,90,90,30), JEAN
 30   CALL FOUR
      CALL SIX
      CALL CHAIN(EDITH)
 90   WRITE(2,901)
 901  FORMAT(T5,'OKAY, YOU ARE OFF')
      CALL VIAQ
      END
```

SUBROUTINE ONE

```
      SUBROUTINE ONE
C
C         THIS SUBROUTINE REQUESTS CONSTANTS OR PARAMETERS
C         FROM POC
C
      DIMENSION GAIL(450)
      COMMON/INSKEL/GAIL,H
      COMMON HOLD1,HOLD2,STEC
      WRITE(2,10)
 10   FORMAT(T5,'ENTER AS REAL NUMBERS'/T5,'HOLD-UP IN FIRST
     1 EFFECT IN
     1BS'/T5,'HOLD-UP IN SECOND EFFECT IN LBS'/T5,'STEAM
     1 ECONOMY'/T5,'C
     2MPUTATION INTERVAL IN SECONDS')
 20   CALL FFINP(1,4,1,HOLD1,1,HOLD2,1,STEC,1,H,KATE)
      IF(KATE) 23,30,23
 23   WRITE(2,231)
 231  FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 20
 30   RETURN
      END
```

SUBROUTINE TWO

```
      SUBROUTINE TWO(JULIA)
C
C        THIS SUBROUTINE OBTAINS PRESENT VALUES OF PROCESS
C        VARIABLES
C        FROM LOOP RECORD TABLE
C
      COMMON HOLD1,HOLD2,STEC,CI,CO,WI,WST
      JULIA=1
C        OBTAIN FEED CONCENTRATION CI
C
 10   CALL GTVLU(323,1,CI,LAURA,1)
      GO TO (20,80,80,80), LAURA
C        OBTAIN PRODUCT CONCENTRATION CO
C
 20   CALL GTVLU(324,1,CO,LAURA,1)
      GO TO (30,80,80,80), LAURA
C        OBTAIN FEED FLOW RATE WI
C
 30   CALL GTVLU(329,1,WI,LAURA,4)
      GO TO (40,80,80,80), LAURA
C        OBTAIN STEAM FLOW RATE WST
C
 40   CALL GTVLU(306,1,WST,LAURA,4)
      GO TO (90,80,80,80), LAURA
 80   WRITE(2,81) LAURA
 81   FORMAT(T5,'GTVLU ERROR CODE',I3/T5,'ENTER 1 TO REPEAT
     1 GTVLU OR 2
     1O GIVE UP')
 82   CALL FFINP(1,1,0,JULIA,KATE)
      IF(KATE) 85,855,85
 85   WRITE(2,851)
 851  FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 82
 855  IF(JULIA) 85,85,86
 86   IF(JULIA-2) 87,87,85
 87   GO TO (10,90), JULIA
 90   RETURN
      END
```

SUBROUTINE THREE

```
      SUBROUTINE THREE(JEAN)
C
C        THIS SUBROUTINE VERIFIES THE MATERIAL BALANCE
C
C
      COMMON HOLD1,HOLD2,STEC,CI,CO,WI,WST,C2
 10   VAPOR=WST*STEC
      C2=CI*WI/(WI-VAPOR/2.0)
      COCAL=CI*WI/(WI-VAPOR)
      WRITE(2,20) CI, CO, WI, WST
 20   FORMAT(T5,'READINGS ARE',T19,'CI=',T22,F7.2,T31,'CO='
     1,T34,F7.2/T1
     1,'WI=',T22,F7.2,T31,'WST=',T35,F6.2)
      WRITE(2,30) C2, COCAL
 30   FORMAT(T5,'CALCULATION YIELDS',T24,'C2=',T27,F7.2,T37
     1,'COCAL=',T4
     1,F7.2)
      WRITE(2,40)
 40   FORMAT(T5,'YOU HAVE THE CHOICE TO'/T5,'1 HAVE READINGS
     1 REPEATED'/
     15,'2 ADJUST READINGS'/T5,'3 CALL IT OFF'/T5,'4
     1 CONTINUE')
 42   CALL FFINP(1,1,0,JEAN,KATE)
      IF(KATE) 43,44,43
 43   WRITE(2,431)
 431  FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 42
 44   IF(JEAN) 43,43,45
 45   IF(JEAN-4) 50,60,43
 50   GO TO (60,52,60,60), JEAN
 52   WRITE(2,521)
 521  FORMAT(T5,'ENTER ADJUSTED READINGS CI, CO, WI, WST')
 53   CALL FFINP(1,4,1,CI,1,CO,1,WI,1,WST,KATE)
      IF(KATE) 54,10,54
 54   WRITE(2,431)
      GO TO 53
 60   RETURN
      END
```

SUBROUTINE FOUR

```
      SUBROUTINE FOUR
C
C        THIS SUBROUTINE CALCULATES CONSTANTS AND
C        INTERMEDIATE QUANTITIES
C
      DIMENSION GAIL(450)
      COMMON/INSKEL/GAIL,H,CST2A,CST2W,CST1B,CST1C,CST9C
     1,CST9W,CST5D
      COMMON/INSKEL/CST5W,CST4E,CST4C,CSTTF,CSTTW,A,B,C,D,E
     1,F
      COMMON HOLD1,HOLD2,STEC,CI,CO,WI,WST,C2
      CST2A=(STEC*WST-2.0*WI)/(2.0*HOLD1)
      CST2W=STEC*C2/(2.0*HOLD1)
      CST1B=(STEC*WST-2.0*WI)/(2.0*HOLD1)
      CST1C=WI/HOLD1
      CST9C=(STEC*WST-2.0*WI)/(2.0*HOLD1)
      CST9W=(CI-C2)/HOLD1
      CST5D=(STEC*WST-WI)/HOLD2
      CST5W=(2.0*STEC*CO-STEC*C2)/(2.0*HOLD2)
      CST4E=(STEC*WST-WI)/HOLD2
      CST4C=(2.0*WI-STEC*WST)/(2.0*HOLD2)
      CSTTF=(STEC*WST-WI)/HOLD2
      CSTTW=(C2-CO)/HOLD2
      A=-WST*CST2W/CST2A
      B=-CI*CST1C/CST1B
      C=-WI*CST9W/CST9C
      E=-C2*CST4C/CST4E
      F=-WI*CSTTW/CSTTF
      D=-WST*CST5W/CST5D
      RETURN

      SUBROUTINE SIX
C        SPARE SUBROUTINE FOR EVENTUAL ADDITIONS TO PROGRAM
C
      DIMENSION GAIL(450),HELEN(19),OLGA(8)
      COMMON/INSKEL/GAIL,HELEN,STONE,BRICK,WALL
      COMMON OLGA,FLOOR,SKY,DUST
      RETURN
      END
```

MAINLINE EDITH

```
      MAINLINE EDITH

C
C          PROGRAM NAME              EDITH
C
C          PROGRAM TYPE              PROCESS PROGRAM,
C                                    MAINLINE CORE LOAD
C
C          ABSTRACT                  PROGRAM INITIALIZES
C                                    INFERENTIAL CONTROLLER
C
C          OPERATION                 PROGRAM IS CONTACTED VIA CALL
C                                    CHAIN FROM DIANE
C
C          SUBROUTINES CALLED        FIVE, REPET, QUEUE, VIAQ
C                                    FFINP, DYTIM, SEVEN, INCHR
C
C
      EXTERNAL COLOR, MVCON, POKUS
      DIMENSION VIOLA(472)
      COMMON/INSKEL/VIOLA, ERICA, IRMA, LOLA, JOSEY, CI, WI,
     1 WST, INES
      CALL SEVEN(LISE,MOLLY)
      GO TO (10,20,30), LISE
 10   CALL FIVE
      CALL REPET
      CALL QUEUE(COLOR,1,0,4,160)
      GO TO 40
 20   CALL REPET
      CALL QUEUE(MVCON,1,0,4,MOLLY)
      GO TO 40
 30   CALL REPET
      CALL QUEUE(POKUS,50,0,4,MOLLY)
 40   CALL VIAQ
      END
```

SUBROUTINE SEVEN

```
      SUBROUTINE SEVEN(LISE,MOLLY)
C        THIS SUBROUTINE SETS FLAGS LOLA AND JOSEY AND
C        SELECTS THE
C        CORE LOAD TO BE CALLED BY TIMER 4
C
      DIMENSION ANNA(473)
      COMMON/INSKEL/ ANNA, IRMA, LOLA, JOSEY, CI, WI, WST,
     1 INES
      LOLA=1
      JOSEY=1
      MOLLY=1500
      WRITE(2,12)
 12   FORMAT(T5,'ENTER NUMBER OF POINTS BY WHICH MODEL
     1 SHOULD LAG PROCE
     1S')
 14   CALL FFINP(1,1,0,INES,KATE)
      IF(KATE) 15,19,15
 15   WRITE(2,35)
      GO TO 14
 19   WRITE(2,20)
 20   FORMAT(T5,'SELECT TIMER CALLED CORE LOAD'/T5,' 1
     1 COLOR',T15,' 2 M
     1CON',T25,' 3 POKUS')
      CALL INCHR(1,LISE,1,3)
      GO TO (60,30,30), LISE
 30   WRITE(2,31)
 31   FORMAT(T5,'ENTER NUMBER OF TIMER CYCLES(INTEGER)')
 32   CALL FFINP(1,1,0,MOLLY,KATE)
      IF(KATE) 33,60,33
 33   WRITE(2,35)
 35   FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 32
 60   RETURN
      END
```

SUBROUTINE FIVE

```
      SUBROUTINE FIVE
C        SUBROUTINE DETERMINES DUTIES OF TIMER CALLED CORE
C        LOAD COLOR
C
      DIMENSION VIOLA(472)
      COMMON/INSKEL/VIOLA, ERICA, IRMA
      WRITE(2,101)
 101  FORMAT(T5,'SHOULD THE INFERENTIAL CONTROLLER'/T5,' 1
     1 RECORD ITS O
     1TPUT ONLY'/T5,' 2 ACTUALLY CONTROL STEAM FLOW')
 15   CALL FFINP(1,1,0,IRMA,KATE)
      IF(KATE) 16,17,16
 16   WRITE(2,161)
 161  FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 15
 17   IF(IRMA) 16,16,18
 18   IF(IRMA-2) 23,23,16
 23   CALL DYTIM(IDA,IRIS)
      WRITE(2,232) IDA, IRIS
 232  FORMAT(T5,'THE TIME IS NOW',I5,I6)
      WRITE(2,234)
 234  FORMAT(T5,'ENTER TIME ELAPSED SINCE ACTIVATION OF DISK
     1 FILE IN MI
     1UTES(REAL)')
 24   CALL FFINP(1,1,1,ERICA,KATE)
      IF(KATE) 245,247,245
 245  WRITE(2,161)
      GO TO 24
 247  GOTO (25,26), IRMA
 25   WRITE(2,251)
 251  FORMAT(T5,'THE INFERENTIAL CONTROLLER WILL BE
     1 ACTIVATED ONE MINUT
     1 FROM NOW AND SEND ITS OUTPUT TO DISK FILE BINGO')
      GO TO 27
 26   WRITE(2,261)
 261  FORMAT(T5,'THE INFERENTIAL CONTROLLER WILL BE
     1 ACTIVATED ONE MINUT
     1 FROM NOW AND SEND ITS OUTPUT'/T5,'TO MEASUREMENT OF
     1 LOOP 0144 AS
     2WELL AS TO DISK FILE BINGO')
 27   RETURN
      END
```

MAINLINE COLOR

```
      MAINLINE COLOR
C
C        PROGRAM NAME              COLOR
C
C        PROGRAM TYPE              PROCESS PROGRAM,
C                                  MAINLINE CORE LOAD
C
C        ABSTRACT                  THIS IS THE INFERENTIAL CON-
C                                  TROLLER. IT READS PROCESS
C                                  INPUTS AND CALCULATES PRO-
C                                  CESS OUTPUTS AT SPECIFIED
C                                  INTERVALS.
C
C        OPERATION                 PROGRAM IS QUEUED BY TIMER 4
C
C        SUBROUTINES CALLED        GREEN, BLUE, WHITE, GTVLU
C                                  PTVLU
C
C
      DIMENSION GAIL(450), HELEN(4)
      COMMON/INSKEL/ GAIL, H, CST2A, CST2W, CST1B, CST1C,
     1 CST9C, CST9W
      COMMON/INSKEL/ CST5D, CST5W, CST4E, CST4C, CSTTF,
     1 CSTTW, A, B, C
      COMMON/INSKEL/ D, E, F, HELEN, IRMA, LOLA, JOSEY, CI,
     1 WI, WST
      COMMON/INSKEL/ INES
      COMMON CO
      DEFINE FILE 25(480,2,U,LOLA)
      GO TO (14,15), JOSEY
   14 CALL GREEN
      GO TO 30
   15 CALL BLUE
      CALL WHITE
   30 CALL VIAQ
      END
```

SUBROUTINE BLUE

```
      SUBROUTINE BLUE
C        THIS SUBROUTINE WILL USE PROCESS MODEL TO CALCULATE
C        PRODUCT
C        CONCENTRATION FOR PRESENT POLL TIME INTERVAL
C
      DIMENSION GAIL(450), HELEN(4), MARIE(3)
      COMMON/INSKEL/ GAIL, H, CST2A, CST2W, CST1B, CST1C,
     1 CST9C, CST9W
      COMMON/INSKEL/ CST5D, CST5W, CST4E, CST4C, CSTTF,
     1 CSTTW, A, B
      COMMON/INSKEL/ C, D, E, F, HELEN, MARIE, CI, WI, WST
      COMMON CO
      XK=CST1C*CI
      XA=CST1B
      POWER=XA*H/60.0
      SECND=XK*(1.0-EXP(POWER))/XA
      B=B*EXP(POWER)-SECND
      XK=CST2W*WST
      XA=CST2A
      POWER=XA*H/60.0
      SECND=XK*(1.0-EXP(POWER))/XA
      A=A*EXP(POWER)-SECND
      XK=CST9W*WI
      XA=CST9C
      POWER=XA*H/60.0
      SECND=XK*(1.0-EXP(POWER))/XA
      C=C*EXP(POWER)-SECND
      C2=A+B+C
      XK=CST4C*C2
      XA=CST4E
      POWER=XA*H/60.0
      SECND=XK*(1.0-EXP(POWER))/XA
      E=E*EXP(POWER)-SECND
      XK=CSTTW*WI
      XA=CSTTF
      POWER=XA*H/60.0
      SECND=XK*(1.0-EXP(POWER))/XA
      F=F*EXP(POWER)-SECND
      XK=CST5W*WST
      XA=CST5D
      POWER=XA*H/60.0
      SECND=XK*(1.0-EXP(POWER))/XA
      D=D*EXP(POWER)-SECND
      CO=E+F+D
      RETURN
      END
```

SUBROUTINE GREEN

```
      SUBROUTINE GREEN

C        THIS SUBROUTINE READS PRESENT VALUES OF PROCESS
C        VARIABLES FROM
C        LOOP RECORD TABLE AND STORES THEM IN INSKEL COMMON
C        FOR LATER USE

      DIMENSION ANNA(473), JOYCE(2)
      COMMON/INSKEL/ ANNA, JOYCE, JOSEY, CI, WI, WST
      CALL GTVLU(323,1,CI,LAURA,1)
      GO TO (2,8,8,8), LAURA
 2    CALL GTVLU(329,1,WI,LAURA,4)
      GO TO (3,8,8,8), LAURA
 3    CALL GTVLU(306,1,WST,LAURA,4)
      GO TO (9,8,8,8), LAURA
 8    WRITE(2,81) LAURA
 81   FORMAT(T5,'GTVLU ERROR CODE',I3/T5,'PREVIOUS LOADS
     1 WILL BE USED')
 9    JOSEY=2
      RETURN


      SUBROUTINE WHITE

C        THIS SUBROUTINE WRITES CALCULATED VALUES OF PRODUCT
C        CONCENTRATION INTO A DISK FILE AND TRANSFERS THEM
C        TO THE DDC LOOP RECORD

      DIMENSION ANNA(473)
      COMMON/INSKEL/ ANNA, IRMA, LOLA, JOSEY, CI, WI, WST,
     1 INES
      COMMON CO
      WRITE(25'LOLA) CO
      GO TO (3,2), IRMA
 2    IF(INES) 23,23,201
 201  LINDA=LOLA-INES
      IF(LINDA) 3,3,21
 21   READ(25'LINDA) CODL
      OUTP=CODL*5.0
      LOLA=LINDA+INES
      GO TO 24
 23   OUTP=CO*5.0
 24   CALL PTVLU(297,3,OUTP,MARY)
      GO TO (3,22,22,22,22,22,22,22,22), MARY
 22   WRITE(2,221) MARY
 221  FORMAT(T5,'PTVLU ERROR CODE',I3,T26,'NO PUTTING THIS
     1 TIME')
 3    JOSEY=1
      RETURN
      END
```

MAINLINE THAT

```
      MAINLINE THAT
C
C        PROGRAM NAME           THAT
C
C        PROGRAM TYPE           PROCESS PROGRAM,
C                               MAINLINE CORE LOAD
C
C        ABSTRACT               THIS PROGRAM PRODUCES A POINT
C                               PLOT OF THE PRODUCT CONCEN-
C                               TRATION CALCULATED BY THE
C                               INFERENTIAL CONTROLLER VERSUS
C                               TIME. THE PLOT OVERLAYS THE
C                               LINE PLOT OF THE MEASURED
C                               CONCENTRATION.
C
C        OPERATION              THE PROGRAM IS CONTACTED VIA
C                               FUNCTION NO. 5 OR
C                               KEYBOARD REQUEST
C
C        SUBROUTINES CALLED     HAZEL, RENEE, FFINP, SCALF
C                               FPLOT, POINT, VIAQ
C
C
      DIMENSION VIOLA(472)
      COMMON/INSKEL/ VIOLA, ERICA, IRMA, LOLA
      COMMON XOLD, YOLD, TOMIN
      DEFINE FILE 25(480,2,U,LOLA)
      CALL HAZEL
      CALL RENEE
      CALL VIAQ
      END
```

SUBROUTINE HAZEL

```
      SUBROUTINE HAZEL

C         THIS SUBROUTINE DETERMINES THE PRESENT PLOTTER PEN
C         POSITION AND
C         MOVES THE PEN TO THE ORIGIN OF THE PROPOSED PLOT

      COMMON XOLD, YOLD, TOMIN
      XMIN=0.0
      WRITE(2,41)
 41   FORMAT(T5,'ENTER INFORMATION FOR SUPERIMPOSED PLOT'/T5
     1,'TOTAL MIN
     1TES, YMAX, YMIN (ALL REAL)')
 42   CALL FFINP(1,3,1,TOMIN,1,YMAX,1,YMIN,KATE)
      IF(KATE) 43,44,43
 43   WRITE(2,431)
 431  FORMAT(T5,'YOU TYPED IN GARBAGE, TRY AGAIN')
      GO TO 42
 44   XOLD=TOMIN+4.0*(TOMIN-XMIN)/8.0
      YOLD=YMIN-2.5*(YMAX-YMIN)/5.0
      XSCAL=8.0/(TOMIN-XMIN)
      YSCAL=5.0/(YMAX-YMIN)
      CALL SCALF(XSCAL,YSCAL,XOLD,YOLD)
      CALL FPLOT(+3,XMIN,YMIN)
      RETURN


      SUBROUTINE RENEE

C         THIS SUBROUTINE PLOTS THE PRODUCT CONCENTRATION
C         CALCULATED BY THE INFERENTIAL CONTROLLER OVER THE
C         PLOT OBTAINED FROM THE RING BUFFER
C

      DIMENSION VIOLA(472)
      COMMON/INSKEL/ VIOLA, ERICA, IRMA, LOLA
      COMMON XOLD, YOLD, TOMIN
      X=ERICA-2.0
      XINCR=3.0*64.0/60.0
      NULLY=LOLA
      DO 59  JOAN=1,NULLY,3
      X=X+XINCR
      READ(25'JOAN) Y
      CALL FPLOT(+3,X,Y)
      CALL FPLOT(-2,X,Y)
      CALL POINT(0)
      IF(X-TOMIN) 59,62,62
 59   CONTINUE
 62   CALL FPLOT(+3,XOLD,YOLD)
      RETURN
      END
```